DAIRY WASTE TREATMENT BY HIGH-RATE TRICKLING FILTRATION, WITH PARTICULAR REFERENCE TO NITROGEN.

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by

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ABSTRACT

The effective disposal of dairy factory waste is becoming increasingly important in New Zealand. Treatment by high-rate trickling filtration is a successful method in use overseas. For New Zealand conditions, a 'roughing' treatment removing 60 - 90% of the BOD of the waste should be adequate. One objective of this research was the development of a filter capable of providing this treatment. Another objective was the resolution of the controversies between the theoretical and empirical performance-prediction relationships available for trickling filtration. Because nitrogen is receiving a greater emphasis as a pollutant, a third objective was the study of nitrogen removal in dairy waste trickling filtration.

The experimental work primarily involved the use of a pilot-scale trickling filter. This was designed using conventional parameters. The filter column was an 18" diameter, 8' long concrete pipe, filled with river stone. An artificial waste compounded from whey and water was fed to the plant at a controlled rate, being diluted with flow from a 25 gallon recirculation tank prior to application to the column. The treated waste overflowed from the recirculation tank and was discharged. The plant was operated at the high organic loading intensities of 1.3 - 2.7 lb BOD/yd^{3}day, and at the high recirculation ratios of 20 - 55 : 1. The levels of BOD and organic, ammoniacal, nitrite and nitrate nitrogen were measured in the feed and settled effluent at different recirculation ratios. Aqueous suspensions of biomass collected from the plant were incubated under aerobic

and anaerobic conditions, in the presence of a variety of carbonaceous and nitrogenous additives. The nitrogen balance of these suspensions was studied.

The plant fulfilled its design function of providing a 'roughing' treatment, as it removed 60 - 85% of the feed BOD. The experimental data did not support the available performance-prediction relationships, and hence the controversies between these relationships were not resolved. The pilot plant performance could be described by the equation

$$Y = 17.778 + 3.079X - 0.0342\ X^2$$

where Y = % removal of applied BOD

X = recirculation ratio

This equation, specific to the pilot plant, predicts an optimum recirculation ratio of 45 : 1, which is considerably higher than the 10 : 1 ratio commonly used. Successful operation of the plant was achieved at BOD : nitrogen ratios in the feed of 21 - 27 : 1, which are higher than the 20 : 1 maximum generally recommended. Despite this high ratio, typically 30% of the feed organic nitrogen was present in the effluent. There was no evidence of nitrification. The nitrogen balance experiments provided evidence of net nitrogen loss from the suspensions, under aerobic conditions. Denitrification under anaerobic conditions followed normal routes.

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INTRODUCTION.

The produce of New Zealand's dairy factories earns approximately twenty percent of its overseas exchange (N.Z. Dept. of Stats. 1969); the same factories produce at least ten percent of its biological industrial waste (Bennett 1969). The effective disposal of this waste is becoming increasingly important for a number of reasons. Firstly, dairy production is increasing and also, in general, is the waste production from both farm and factory. Because much of this material is water-carried, the greater waste production is increasing the loading of pollutants on waterways in dairying areas. Secondly, amalgamation of small factories into larger units is also causing higher pollution levels in streams which were possibly able to cope with the discharge from the smaller units. Thirdly, there is the recent upsurge in public interest in the pollution and conservation of the environment. This should tend to make public authorities more stringent in their enforcement of pollution regulations, and effective waste treatment, or disposal, more necessary.

Dairy factory waste can be disposed of or treated by a large number of methods, as outlined in Table 1. Disposal may be by dilution with large volumes of water, such as in rivers, lakes and the sea. Another method is spray irrigation, onto pastures or forest and scrub. Treatment may be by any of the methods used for domestic waste, and includes chemical precipitation, incorporation into domestic sewage for municipal treatment, aerobic processes such as the various forms of trickling filtration, activated sludge and oxidation pond, and

Table 1.

Dairy Waste Disposal and Treatment.

Disposal Methods	Comments
Dilution into a large volume of water e.g. sea, lake, river.	Permissible only where effect on water is slight.
Spray irrigation on pastures, forest and scrub.	Can be beneficial if properly controlled according to conditions e.g. soil, weather.
Treatment Methods	
Chemical Precipitation	Can be successful, but high cost.
Incorporation into municipal treatment	Worthwhile if charges are reasonable.
Activated sludge process	Quite successful, requires little space. Sensitive to load variation.
Trickling filtration	Generally the most successful treatment method
Oxidation Pond	Large area required. Not common.
Anaerobic Treatment	Good for excess sludge from trickling filtration and activated sludge. Large volume and close control of raw waste required.

anaerobic digestion. Trickling filtration is the most commonly favoured method of treatment, its virtues including greater stability under varying load than its closest rival, the activated sludge process.

The treatment of dairy waste by trickling filtration was chosen as the basis of this study. The investigation considered three main topics:

(a) The treatment of a compounded dairy waste by means of trickling filters operating at hydraulic and organic loadings greater than used in normal practice.

(b) A study of the many theoretical and empirical relationships for filter design and performance and the applicability of experimental data to these predictions.

(c) A study of nitrogen relationships in this type of trickling filtration.

Other topics briefly considered were oxygen transfer rates in the experimental units, the bacterial composition of the growth on the units, and the digestion of excess growth by anaerobic means. The literature relevant to each major section of the study is considered at the beginning of its respective chapter, although inevitably there is some overlap.

CHAPTER 1.

DAIRY WASTE TREATMENT BY TRICKLING FILTRATION.

INTRODUCTION.

This chapter will describe the design and building of the experimental plants, their general day-to-day operation, and an assessment of their ability to treat dairy factory waste. For this purpose it is divided in the following manner:

Section A: The collection of general background information and design data on trickling filtration as a means of treatment for both domestic and industrial waste.

Section B: The collection of information on dairy waste treatment, in particular by trickling filtration.

Section C: The co-ordination of information from A and B into the design of pilot and laboratory scale plants, and the building of these plants and their operation.

Section D: The comparison of the performance of these plants with those cited in the literature, and a discussion of the suitability of the experimental plants for dairy waste treatment.

SECTION A: GENERAL BACKGROUND INFORMATION AND DESIGN DATA.

Organic waste material causes pollution of waterways primarily because of the oxygen requirement of the material for decomposition. The oxygen demand removes oxygen from the water; oxygen is thus less available for the normal aquatic flora and fauna. If the oxygen level falls below the limit for any particular species, that species will move out of the polluted zone, if it is able to, or it will die, its decomposition in turn increasing the oxygen deficit.

The basis of the treatment of organic waste material is to remove the dissolved or suspended organic matter from its carrier, water, by converting the solids into a recoverable form, such as settleable solids, or into a less harmful form, such as the metabolic products carbon dioxide and water. The two major aerobic processes for the biological conversion of solids in this way are trickling filtration and activated sludge digestion. Both processes do this by biological flocculation and precipitation by means of biomasses grown in the treatment plants, but in activated sludge units the biomass is in suspension in the bulk waste liquid during aeration, whereas in trickling filtration plants the biomass is supported on a fixed solid support. In both cases excess biomass is removed from the system, normally by sedimentation.

Trickling filters can therefore be considered basically as static-bed biological reactors over which the reactant material flows. McKinney (1962) described them more crudely as "a pile of rocks over which sewage or organic wastes slowly trickle". Today there are many different forms of trickling

filter, all of which fit these basic descriptions. The filters may also be described by a variety of names, including sprinkling filter, percolating filter, biological filter, biofilter and bacteria bed. The term "trickling filter" is the most common, although strictly speaking it is inaccurate as the process does not involve filtration in the conventional sense. According to Stanbridge (1954), trickling filtration developed from the centuries-old practice of land application of sewage in 1887, and so it is not surprising that filters have evolved into many different forms under different names from that period to the present.

Design Parameters

There are four parameters of importance in the classification of filters into their various types; these same parameters are of importance in design and performance assessment. The parameters are organic loading intensity, hydraulic loading intensity, recirculation ratio and efficiency. The definition of these parameters is complicated by the fact that there is controversy over the unit measure of a filter - basically whether it should be expressed in terms of horizontal area occupied by the filter or on volume. The tendency in the U.S.A. is to use area as the basis, in contrast to the British tendency to use volume. The volume basis will be used in this thesis, as it gives more meaningful measure of filter size.

Organic Loading Intensity is an expression of the time rate of application of organic material per unit filter measure. The organic load is normally expressed in terms of biochemical

or biological oxygen demand i.e. BOD. The BOD is measured by a standard test which determines the oxygen demand of sewage, sewage plant effluents, polluted waters or industrial wastes exerted by (a) carbonaceous organic material useable as a source of food by aerobic organisms (b) oxidisable nitrogen derived from nitrite, ammonia and organic nitrogen compounds which serve as food for specific bacteria, according to the American Public Health Association et al (1960). The test usually measures the demand over a period of five days, the samples being incubated at 20°C. Hence arises the term five day BOD. The combination of organic load measurement and filter unit measure to provide an expression of organic loading intensity results in terms such as lb BOD/cubic yard day, lb BOD/acre foot day, lb BOD/1000 cubic feet day, lb BOD/acre day and kgm BOD/cubic metre day. The term lb BOD/cubic yard day has been adopted for the bulk of this thesis.

Hydraulic Loading Intensity is an expression of the volume rate of application of waste per unit of filter measure. It may be used to express the rate of application of raw waste to the filter or, where recirculation is practised, of raw waste diluted with filter effluent. The units used include both Imperial and U.S. gallons. Some of the terms are million gallons/acre day, gallons/acre foot day, gallons/1000 cubic feet day, gallons/cubic yard day and cubic metres/cubic metre day. In keeping with the expression for organic loading intensity, the unit Imperial gallon/cubic yard day was chosen.

Recirculation Ratio gives an indication of the degree of dilution of raw waste being fed to the filter with effluent from

the filter. Being a ratio it is dimensionless and is defined as the difference between the diluted and undiluted hydraulic loading intensity divided by the undiluted hydraulic loading intensity i.e.

$$\frac{\text{Recirculated flow rate - Raw waste flow rate}}{\text{Raw waste flow rate}}$$

The efficiency of a trickling filter is normally expressed as the percentage of the applied BOD that is removed during treatment.

As already stated, these four parameters form the basis of the broad classification of trickling filters. One such classification is that given by McKinney (1962). Table 2 is based on this.

Table 2.

A Classification of Trickling Filters

Filter Type	Hydraulic Loading (recirc. & raw)		Organic Loading	Recirc. Ratio	Effic.
	mgad(U.S.)	Ig/yd^{3}day	lb/yd^{3}day		%
Low Rate	2-4	172-345	0.27-0.68	Nil	85-90+
High Rate	10-40	860-3450	1-2	0.5-10:1	65-75
Super Rate	100(to 400)	8618	2.5-10	10-50:1	65(-95)

It is important to note that the hydraulic loadings were not expressed by McKinney on the volume basis of Ig/yd^{3}day; the conversion to this form was an arbitrary one based on a filter depth of six feet.

To aid in the design of experimental plants data was collected from a large number of sources on recommendations for the design of trickling filters, in general for domestic sewage

treatment. These are presented in Table 3. Again, an arbitrary depth of six feet was chosen to convert flow rates in mgad to lg/yd^{3}day where necessary.

Table 3.

Design Parameters for Filters

Author	Filter Type	Flow Mgad	Flow lg/yd^3 day	Recir. Ratio	lb/yd^{3}day Organic Load	Depth feet
Bryan & Moeller (1963)	Super Rate				2.0	
Bruce (1969)	Low		81	Nil	0.17	6
Eckenfelder & O'Connor (1961)	Low	2-6	172-517	Nil	0.15-0.77	6
	High	10-100	860-8600			6
Fair & Geyer (1954)	Low	2-6	172-517	Nil	0.15-0.77	6-10
	High	15-30	1293-2585		1.03-4.6	6-10
Fair,Geyer & Okun (1968)	Low	1.1-4.4	95-379	Nil	0.13-0.67	5-8
	High	8.7-44	750-3790	0.5-10	0.67-0.81	3-8
Gurnham (1955)	Low	1-4	86-345	Nil	0.12-0.25	6-8
	High	10-30	860-2585	5	1.1-1.9	6-8
Hanumanulu (1969)	High		340	1.5	1.9	12
Heukelekian (1945)	Low	3.8	327	Nil	0.3-0.4	6.5
	Medium	6.3	543	1.5	1.5	3
Lesperance (1968)	High	10-40	860-3450		0.015-3.0	
	Super	200-400	17,200-34,500		5-10	
Wittmer (1948)	Low			Nil	0.4	5.5

Humus Settling Tank

An integral part of any trickling filter is the humus settling tank. This is used for removing from the column effluent surplus growth of biomass which is flushed off the column by the flow of liquid through it. Frequently also the raw waste being fed to the filter passes through a settling tank. There are many possible arrangements of filters and settling tanks; Fig. 1 below shows examples of these. Although settling tanks vary greatly in design, theoretical residence times are a frequently quoted design parameter. Some typical values are shown in Table 4.

Table 4.

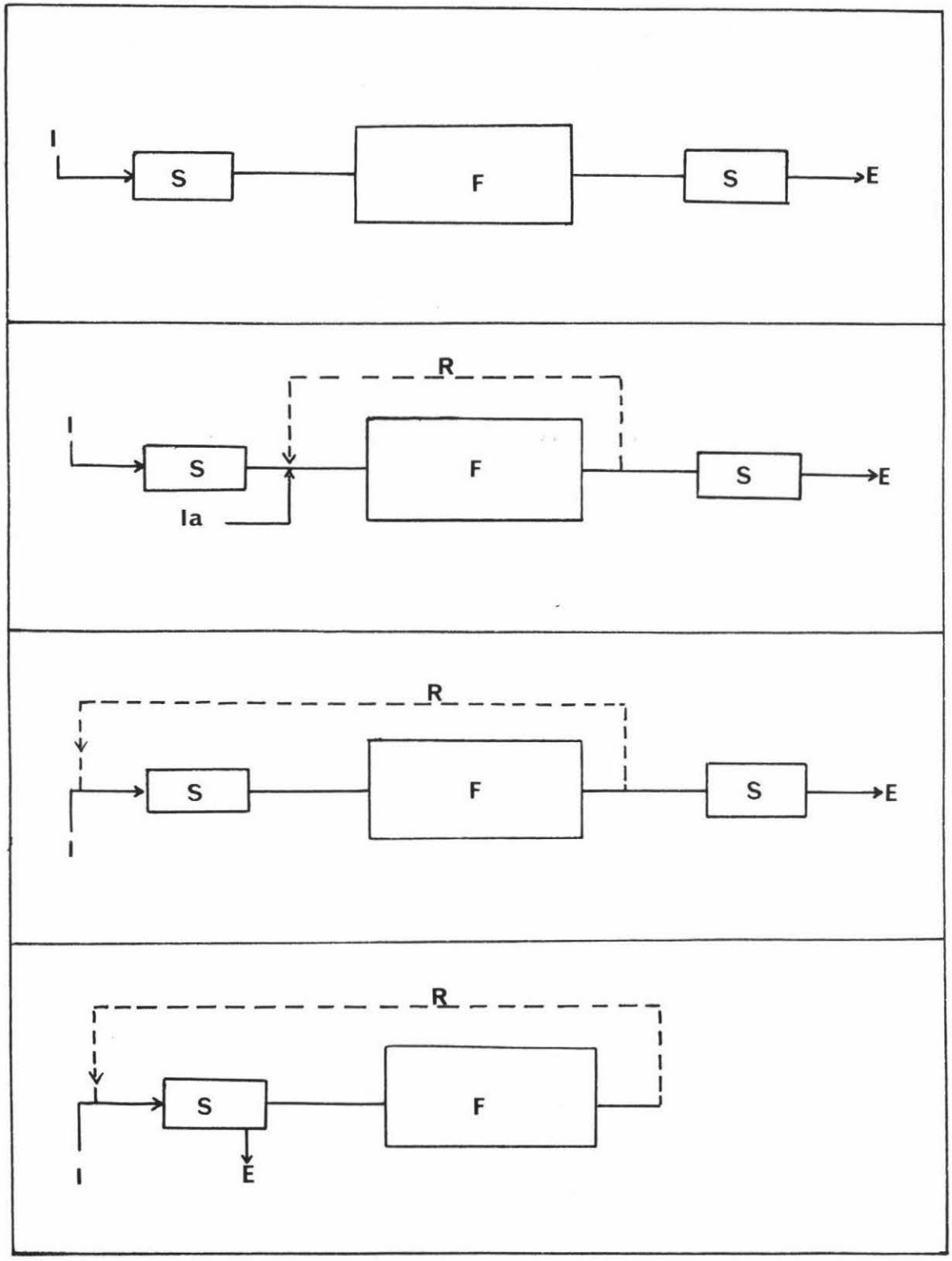
Settling Times

Author	Settling Time (Holding time)
Dairy Research Inst. Prague (1966)	2.6 hours
Fair, Geyer, Okun (1968)	1-2 hours for 10' depth
Gurnham (1955)	1-2 hours
Greeley (1948)	2.5, 1.6, 2.5, 1.25 hr (Different plants)
Silvester (1959)	3 hour (upward flow pyramid)

Filter Packing Media

A very important feature of trickling filter design is the choice of packing material to support the microbial growth. The table below (Table 5) summarises a number of opinions on the requirements of the packing.

Fig.1 PLANT LAYOUTS



I	Influent	E	Effluent	S	Settling
Ia	Influent alternative	F	Filtration	R	Recirculation

Table 5.

Filter Media Characteristics

Author	Media size (max. dimension)	Materials	Properties Required
Gurnham (1955)	2-3"	Broken brick slag, clinker, anthracite, trap rock, granite.	Spherical, strong enough to form a bed. Chemical resistant. Readily available. Cheap. Large free cross-section. High Liquid retention.
Fair & Geyer (1954)	1½-3"	Crushed stone, hard coal, coke, slag, wood, ceramics, plastics	Small enough for large surface, but not small enough to clog.
Bruce (1969)	1½-2"	Clinker, broken rock, gravel, slag.	2" size is a compromise between surface area and non-clogging.
Steel (1960) Fair, Geyer & Okun (1968)	1½-3½"	Crushed stone, slag, gravel.	Must withstand physical stresses. Uniform in principal dimensions, 95% passing 4" square screen, but held on 2½" screen.

Size of Pilot Plants

An important aspect of any pilot-scale plant is that the results obtained from this plant can be meaningfully used for the design of full-scale plants. For this to be possible, the size of the pilot plant must be sufficiently large to prevent distortion of the results beyond what can be allowed for by the use of scale factors. Wilson (1959) made some valuable comments on the size requirements of pilot-scale trickling filters. Noting that the purification capacity of a filter is a function of the available active surface, he commented that the small size media frequently used in pilot plant work often gave misleading results because of the relatively high surface area of small media, but that this was probably compensated for by the more frequent clogging of the small-media filters. He claimed a packing diameter of greater than one-eighth the column diameter caused short-circuiting of liquid down the column walls, and that the smallest packing representative of normal plant usage was two-and-a-half inches; on this basis, the minimum pilot plant diameter should be eighteen inches approximately. Wilson has apparently made an error, possibly deliberately, in this filter diameter, as the product of eight and two-and-a-half is twenty, not eighteen. The information he gave is never the less valuable for use in pilot plant design.

SECTION B: DAIRY WASTE TREATMENT, BY TRICKLING FILTRATION IN PARTICULAR.

The following discussion highlights some of the major developments in the application of trickling filtration to the treatment of dairy factory waste. Experimental work is given especial emphasis.

Dairy wastes have been treated by the trickling filtration process for at least sixty years. Kimberly (1909) described experimental work on dairy waste treatment, using baffled sedimentation units and sand bed filters. The effluent so treated was generally unsatisfactory. Twelve foot deep clinker filters following chemical precipitation were recommended for dairy waste treatment by Kershaw (1914). A unique wood lath type of filter was used by Levine (1929). He found that lactose was readily oxidised on this and that a greater depth of filter was necessary for the equivalent treatment of cheese factory wastes than normal dairy wastes, the former having a lower nitrogen content. Recirculation was being used for dairy wastes from an early date, as described by Elridge (1939).

A classic work on dairy waste treatment was published by the Department of Scientific and Industrial Research (U.K.) (1941). This was the result of several years of experimentation carried out by the U.K. Water Pollution Research Laboratory. The experiments covered all aspects of dairy wastes and their treatment, on small and large scale. The conclusions and recommendations decided on were:

(a) That the waste waters from milk collecting and

distributing depots and from cheese factories could be satisfactorily treated in percolating filters and by the activated sludge process.

(b) Of these processes, treatment in series on two percolating filters with periodic change in order was the most economical and convenient. (This process soon became known as alternating double filtration, or A.D.F.)

(c) That the plant should contain a storage and balancing tank of sufficient size to allow the filters to be supplied at a constant rate during twenty-four hours. The crude waste BOD recommended was not above 300 ppm, and, if above, dilution with water or final effluent to this level was considered desirable. The maximum hydraulic loading including dilution if necessary was not above 320 g/yd^{3}day for the primary filter.

(d) That there should be sedimentation between and after the two filters, and the hydraulic loading to the second filter should be the same as to the first. The order of the filters should be reversed every two weeks.

These recommendations provided the basis for the design of a large number of trickling filter installations in the United Kingdom.

Southgate (1948) discussed the treatment of milk wastes by the addition of coagulants, followed by sedimentation and trickling filtration, and concluded that both this method and alternating double filtration were processes capable of producing satisfactory effluents.

The fact that milk wastes are more amenable to treatment than domestic sewage was emphasised by Trebler and Harding (1948).

They described trickling filtration plants in the U.S.A. capable of achieving the very high rate of removal of 9.5 lb BOD/yd^3day. Sedimentation prior to recirculation was generally practised.

Gloyna (1950) reported on a laboratory study of dairy waste treatment, using an eighteen inch diameter filter capable of depth variation of one to seven feet. He used a compounded waste composed of a skim milk and water mixture. He concluded (a) that the rate of extraction (i.e. organic load removed/unit depth) increased with an increase in hydraulic loading, for all filter depths. (b) That the average rate of extraction for all filter depths was about the same when the hydraulic loading (diluted) was kept constant and different recirculation ratios were employed.

Harding (1952) commented that for twenty years trickling filters had been recognised as the most practical method of milk waste disposal, and that the trend was to the use of high rates of recirculation (10:1 plus), high flow rates (20-30 mgad plus), shallow filters (3.5 - 6 ft), large media (3 - 3.5 inch diameter), and storage capacity for twenty-four hour load equalisation. An important comment was that the trend was to give at least an hour's retention and settling to that liquid discharged as final effluent from the filter, and little or no settling to the liquid being recirculated. Common organic loadings quoted were 1-2 lb/yd^3day, with BOD reductions of 90 - 97% being possible.

Aeration tanks following high rate trickling filters were suggested by Trebler and Harding (1955) as providing the

best treatment for dairy waste.

Thomas (1958), in a review of dairy waste treatment, commented that low rate filters were unsatisfactory because the rate of flow for a given BOD load was much lower than for domestic sewage. The corresponding decrease in flushing action of this flow caused clogging of the filter.

A discussion of general methods of dairy waste treatment by Wheatland (1959) favoured alternating double filtration at an organic loading of 0.48 lb BOD/yd^3day based on the combined volume of the filters. An interesting comment was that one volume of dairy waste could be recirculated against a ten foot head 170 times before pumping costs equalled aeration costs to treat that waste volume by the activated sludge process.

A novel pilot scale trickling filter was described by Ingram (1961). It was a two-stage unit with recirculation, each stage being ten feet deep, with air blown in at three foot intervals. Very high organic loadings of 18.6 lb BOD/yd^3day with 65 - 75% BOD removal were claimed to be possible.

Svoboda (1964) described the use of single-stage fermentation followed by deep trickling filtration and found loading could not be as high as for domestic waste, but that levels of 1 - 1.3 lb BOD/yd^3day were possible. A further report by Svoboda (1966) stated that the tower filters used operated successfully even at ambient temperatures as low as -14.5°C.

Fisher (1958) presented a review of the treatment and disposal of dairy wastes and stated that the best comparison of treatment methods was obtained by pilot scale testing, but that trickling filtration generally proved the best.

Popovich (1969) claimed that the retention period of the dairy waste on a filter was highly significant in the interpretation of the treatment kinetics.

The foregoing comments and many others on the design and use of trickling filters for dairy waste treatment are summarised in Table 6.

Compounded Waste for Use in Experimental Studies

For meaningful results to be obtained from experimental dairy waste treatment, the waste used in the experiments should be representative of typical dairy waste. The considerable range in BOD values of dairy wastes is shown in Table 6. However, not only does BOD vary greatly, but also the composition of the waste, this being dependent on the products manufactured and on the general manufacturing procedure. Generally, dairy wastes are dilutions of the raw materials used and the products made. McDowall and Thomas (1961) presented the data in Table 7 for undiluted milk and its liquid derivatives. The strength of the waste depends on the degree of leakage of these materials into the waste stream. Actual data on waste production on the basis of milk processed or product manufactured are not frequently found in the literature; some are presented in Table 8. The different types of compounded wastes that have been used in experimental dairy waste units are presented in Table 9.

Table 6.

Performance of Dairy Waste Trickling Filters.

Author	No. of Stages	Hydraulic Loading (recirc. & raw) Ig/yd^3day	Organic Loading lb BOD/yd^3day	Wt. BOD removed lb/yd^3day	Efficiency % BOD removed	Raw Waste Strength ppm BOD	Recirc. Ratio	Depth
Davy (1952)	1	1360	1.64	0.85	52 (not settled)	760	5.4:1	
D.R.I. (Prague) (1966)	1	1700	9.4	5.82	62	1400	1.54:1	
D.S.I.R. (U.K.) (1941)	2 ADF	320 each 160 comb	0.36 comb.	0.35	97	Adj. to 300	Variable	4'5"
Eckenfelder & O'Connor (1961)	1	1800	1.42	1.31	92	1160	13.5:1	4'
Elridge (1939)	1	128	0.508	0.46	90	537	0.35:1	7'
Fraser (1968)	1	1545	1.78	1.32	74	1150	10:1	6'
Gloyna (1950)	1		1 - 5		Varied		Variable	1'-7'
Gurnham (1955)	1	1400	3.2	2.88	90	2500	10:1	6-10'
Harding (1952)	1	1680-2500*	1-2*	1.41	90-97	600	5:1-10:1	3.5-6'
Hatch,Bass (1939)	1	100	0.1	0.10	99	1291	12:1	
Ingram (1961)	2	15,500 each	18.6 comb	13.0	70	546	4.5:1	10'
Morgan & Baumann (1957)	1	277	1.135	1.1	97	3720	7.75:1	
	1	133	0.67	0.64	96	1350	3.11:1	
Muers (1968)	1	500	5.0	3.5	60-80	2000	1:1	6'
Trebler & Harding (1948)	2	2480 comb	1.93 comb	1.74	90	855	10:1	3'6"
Wilson (1946)	2	1660 comb	1.9 comb	1.56	82	1030	8:1	
Wittmer (1948)	1		0.4	0.30	74			5'5"
Zack (1953)	2	480* each	0.72 each	0.68	94	225-300	1:1 ea.	
Lab. Plant	1	58,800	1.04	1.01	97.3	612	344.5:1	4'
		58,800	2.13	2.10	98.8	1237	344.5:1	4'
Pilot Plant	1	3471	2.64	1.69	64	1620	20:1	8'
		4972	3.38	2.73	81	2078	30.5:1	8'
		5035	1.99	1.57	79	1172	28.6:1	8'
		6595	1.48	1.27	87	870	38.8:1	8'
		8915	1.84	1.53	83	1080	51.5:1	8'
		9420	1.72	1.46	85	1012	54.5:1	8'

* Depth of 6' chosen to convert mgad - Ig/yd^3day.

Table 7

BOD of Dairy Products.

Material	5 day BOD ppm.
Whole Milk	102,000
Skim Milk	73,000
Buttermilk	64,000
Whey	32,000

Table 8.

Dairy Wastes.

Author	Waste Type	Waste Volume	Waste Strength
Elridge (1942)	Skim Milk plant waste	1.2 Imp.gall per lb. powder	800 ppm BOD
McDowall & Thomas (1961)	Casein plant washings	Twice volume of raw milk	Equivalent of 10% raw milk vol. as whey i.e. 1600 ppm BOD.
Veale (1941)	Cheese	0.26-0.84 I.gall per lb cheese	
	Butter	0.71-1.35 I.gall per lb butter	
	Lactic casein	3.37 I. gall per lb casein	
Wheatland (1959)	Milk collecting plant	0.15-1.5 x vol. of milk	With 0.5% milk loss 550 ppm BOD
	Butter or cheese	1-2 x vol. of milk	If buttermilk and whey used, 1500-3000 ppm BOD.

Table 9.

Compounded Wastes.

Author	Compounded Waste	Strength
Adamse (1966,1968)	3 parts Hatmaker whey powder 1 part skim milk powder 1 gm mixture/litre	660 ppm BOD
D.S.I.R.(U.K.) (1941)	Dilutions of whey and milk to various strengths	
Ingram	Cheese whey diluted with water	546
Gloyna (1950)	Pasteurised and boiled skim milk diluted with water.	

SECTION C: THE DESIGN, CONSTRUCTION AND OPERATION OF THE EXPERIMENTAL UNITS.

Two experimental units were built, the first a small laboratory scale plant and the second a larger pilot scale unit. The two plants were designed for different purposes and hence differed considerably. Both underwent evolution as more design information became available and as their operation demanded.

Laboratory Scale Plant

This was built within two weeks of the start of work on the Masterate degree course and was operated for a period of eighteen months. Its purpose was threefold:

(a) To provide a source of biomass stabilised to dairy waste as seed for a larger experimental unit.

(b) To provide information on the general operation of trickling filters and on problems likely to be encountered in the operation of a larger plant.

(c) To provide information on trickling filtration using very high rates of recirculation, which are easily obtainable on a small-scale plant.

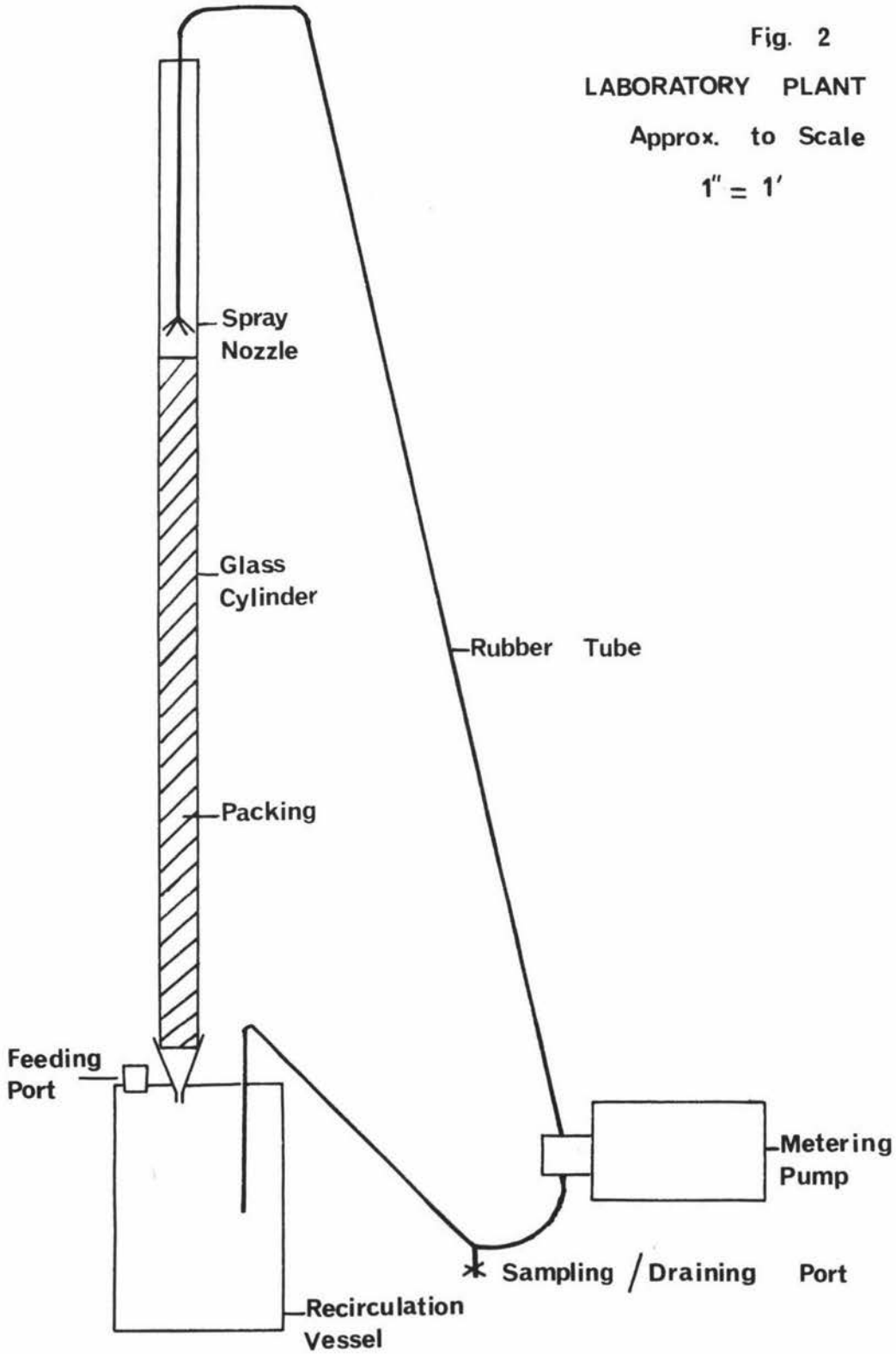
The plant was built prior to the accumulation of the bulk of the information presented in this chapter; this is the reason for some of its unconventional design. The basic labout is shown in Figure 2 and Plates 1 and 2. It was a two-inch internal diameter glass column of five foot length, filled to the four-foot mark with foamed polystyrene in the form of cylindrical pieces approximately 2" x ½". The column was fitted with a plastic spray nozzle immediately above the packing, and this

Fig. 2

LABORATORY PLANT LAYOUT

Approx. to Scale

1" = 1'



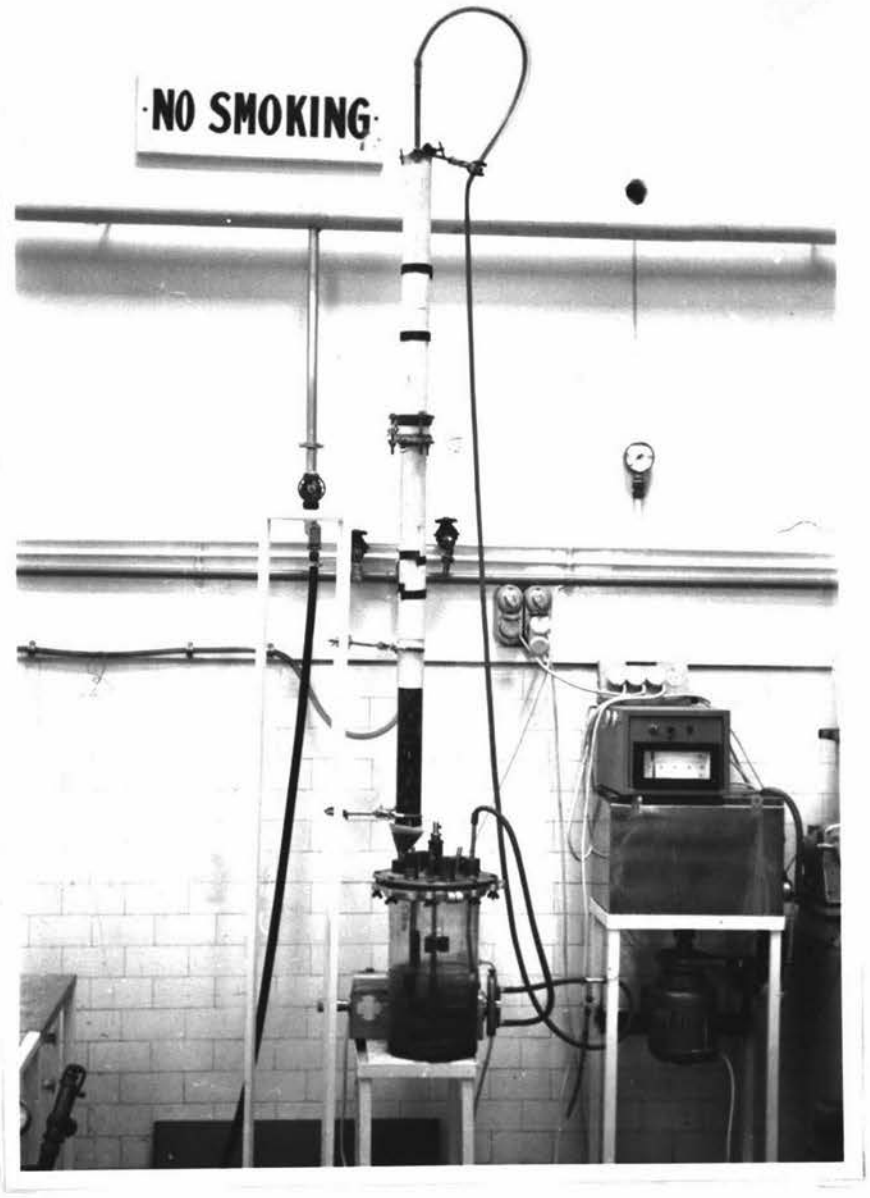


PLATE 1.

LABORATORY SCALE PLANT.

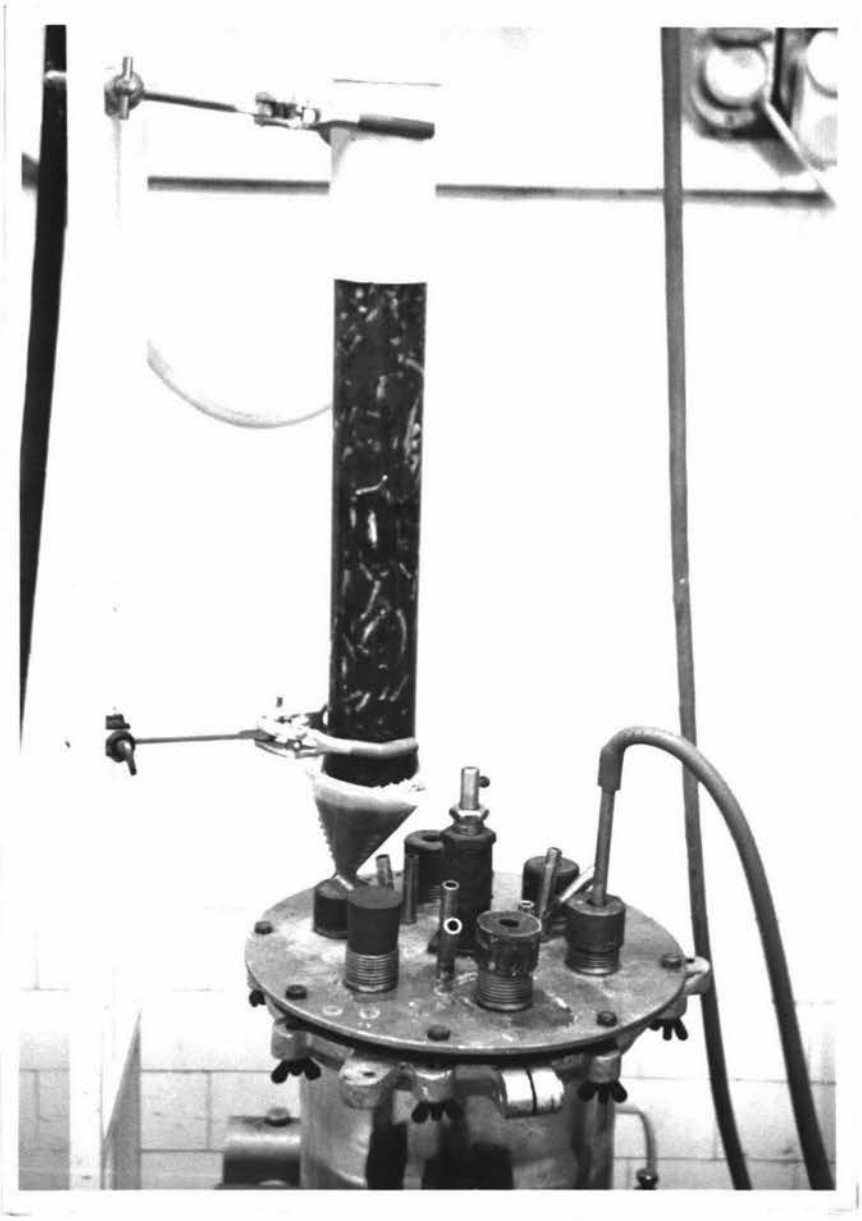


PLATE 2.

FLORA AND PACKING OF LABORATORY PLANT.

nozzle was connected by rubber tubing to the delivery side of a metering pump. The suction line of this pump led to the bottom of a ten-litre cylindrical glass vessel, into which flow from the base of the column was designed to enter. The plant was designed as a batch unit, and for this purpose was fitted with feeding and drainage points. The column was fitted with removable cardboard covers to prevent algal growth.

Plant Operation. The plant was commissioned by filling the glass recirculation vessel to the five-litre mark with a mixture of three parts whey powder/one part skim milk powder at one gram per litre to give a theoretical BOD (Adamse, 1966, 1968) of 660 ppm. The pump was set at 600 cc/minute; this gave an hydraulic loading of 58,880 Ig/yd^{3}day.

The daily procedure of plant operation was to drain off half the volume of liquid i.e. 2½ litres, and to replace this with 2½ litres of compounded dairy waste of the formula above. A daily sample was drawn and stored for later determination of pH and lactose level. (Appendix 1). A daily log was kept of plant operation, and changes to the plant were noted.

Plant Performance. Within three days of the start of operation, a buff-coloured growth became noticeable on the column packing. The lactose level in the daily sample dropped from an initial level of 444 ppm to zero within fourteen days of start up of the column and thereafter was generally zero throughout the plant operation; on only eight occasions was it found to be above zero. In view of the high degree of lactose removal, the feed to the plant in terms of organic load was increased twice.

One month after the start of operation it was increased to 1.5 gm of powder mix per litre, and after approximately one year a mixture of 1.7 gm whey powder per litre of water was used. (This was a dilution of the whey powder concentrate being fed to the pilot plant at this time.) Biochemical oxygen demand removal (Appendix 1) was measured in the last six months of plant operation; these results are summarised in Table 10. Rates of lactose removal were estimated from time to time. These are dealt with more fully in Chapter 4.

Very little difficulty was experienced in the operation of this plant throughout the eighteen months. Little excess flora was produced, the recirculation vessel being cleaned only twice. The only problem was the occasional blockage of the column by excessively heavy growth. This was not unexpected, in view of the small size of packing used, with little void space.

Pilot Scale Plant

Much more information had been collected when the pilot plant was designed than for the laboratory plant, but the plant still underwent evolution as more information became available and as conditions demanded. The plant was designed with three main objectives:

(a) To assess the suitability of a filter operating at higher-than-normal hydraulic and organic loadings for the treatment of dairy wastes.

(b) To provide information on the relationships between hydraulic loading rate, organic loading rate and BOD removal.

(c) To provide information on nitrogen removal in

trickling filters, and to provide a source of organisms for detailed study of nitrogen relationships.

To fulfil these objectives, the filter had to be reasonably close to possible commercial design for valid comparisons with other plants, had to have variability of hydraulic and organic loading and had to have facility for the control of the variables affecting the system.

The decision to design the filter for operation at higher hydraulic and organic loadings than normal was made for two main reasons. Firstly, little work had been done on the treatment of dairy waste by trickling filtration in this region of loadings. Secondly, it was decided that for New Zealand conditions with relatively mild requirements compared with the U.K. or U.S.A. for the degree of purity of waste water, the most suitable type of treatment would be a 'roughing' one similar to that given by super-rate filters. This implies a high rate of organic load removal per unit of filter size, but a higher level of BOD than with lower-rate filters. The high cost of synthetic, plastic media for super-rate filters makes their adoption in New Zealand unlikely, and it was hoped that a similar type of treatment could be obtained by operating a rock media filter at higher loading rates.

Because of the limited funds available for this research, the design and construction of the plant was somewhat a compromise between what was desired and what equipment was available. After a period of evolution, however, a satisfactory plant arrangement was developed.

Column Size. The first item to be decided on was the

size of column. On the basis of the comments of Wilson (1959), discussed in Section A, a column diameter minimum of eighteen inches was chosen. To aid in the simulation of super-rate columns, a greater depth than the conventional six feet was desired. Cheaply available was a slightly damaged concrete pipe of eighteen inches internal diameter and eight and a quarter feet length. This was chosen on the basis of the filter.

Filter Packing Media. Bearing in mind the information in Table 5, Section A, an investigation was made of locally available supplies of media. The only material suitable found was river stone. Stones of approximately spherical shape were chosen, and these were roughly sized on a 2½" square holed screen, stones passing through being discarded, as were stones approximately 3½" or more in largest dimension.

Settling Tank. Initially it was proposed to allow sedimentation of both effluent being discharged from the system and that being recirculated to the filter. (The paper by Harding (1952) recommending settling for the discharged flow only had not been read at this stage.) It was envisaged that the maximum hydraulic loading including recirculation would not exceed five times the highest loading for conventional plants found in the literature. Inspection of the data in Table 6 showed this was 2480 Ig/yd^{3}day, as given by Trebler and Harding (1948). On the basis of a column volume calculated at 0.524 cubic yards, the maximum volume to be handled by the settling tank worked out at 6550 Ig/day, or 273 gallons/hour. From the data presented in Section A, Table 4, one hour holding time was chosen as being

reasonable. Available was a stainless steel tank with a 300 gallon capacity and a gently sloping bottom that had originally been used for milk cooling. This was chosen as the settling tank, and was fitted with an overflow pipe at the 245 gallon mark.

Feed System. To simulate true industrial conditions, the plant was to be run on a continuous feed/continuous outflow basis. For the commissioning period a whey/water mixture of approximately 1000 ppm BOD was proposed, on the basis of the data presented in Table 6 and in Table 8. Cheese whey was available free from the nearby New Zealand Dairy Research Institute. Accurate metering of both the whey and water flow and also adequate mixing of the two streams was required. An accurate metering pump of suitable flow range was available for the whey feed, and it was proposed to use a constant head water tank discharging through an orifice to supply water at a constant rate. The two streams were designed to discharge at right-angles to each other into a plastic cylindrical vessel of approximately 100 cc capacity. The orifice used after some trial and error was from a hypodermic syringe.

It soon became apparent in the design stage that segregation of the influent and the effluent of the plant was a problem with the equipment available. There were a number of possible arrangements. One was a mixing tank separate from the settling tank into which was fed the raw feed and the partially treated effluent for recirculation. This arrangement required two pumps - one from the settling tank to the mixing tank and another from this to the top of the column. Another was arranging the feed

to enter directly to the top of the column. A third was to have the feed enter the settling vessel, but in some way be segregated from the final discharged effluent. The first possibility was eliminated because at this design stage only one pump was available. The second was eliminated because it would have been very difficult physically to arrange the water feed tank at the top of the column. The remaining method was developed by fitting the suction line of the pump supplying the flow to the filter into a cylindrical vessel 8" diameter and 15" depth. This was punched with holes and placed in the settling tank so the top was just above the liquid surface. The holes created sufficient resistance to the liquid flow for the liquid level in this suction vessel to be below that in the surrounding settling tank, thus minimising back-flow into the settling tank and hence to the plant outlet. The feed line led into the pump suction vessel and was thus segregated from the outflow.

It was decided on the basis of the literature search that the organic loading for start-up and general maintenance of the column would be 3 lb BOD/yd^3day. This was higher than the values in common use, but the aim was to run the column at higher loadings. For the 0.524 cubic yard column this converted to 1.57 lb BOD/day. On the basis of a whey BOD of 32,000 ppm (McDowall & Thomas, 1961) this worked out at 4.908 Ig whey/day, and the metering pump was adjusted to this. For the water feed, this corresponded to a water flow rate of 152 Ig/day, and the orifice in use at this time, a 'Saunders' valve, was adjusted to this.

Recirculation Pump. At the time of building the plant, only one pump was available. This was a rubber diaphragm type

normally used in milking sheds, and gave a flow rate of one gallon per minute. This corresponded to an hydraulic loading of 2750 lg/yd^{3}day, with a recirculation ratio of 8.2:1 on the above feeding rate. This was rather lower than desired, but it was thought suitable for commissioning the plant.

Plant Commissioning and Operation. The plant was operated for a period of eight months, and during that period underwent considerable evolution until a satisfactory arrangement was developed. The major changes are shown by a comparison of Figs. 3 and 4, the original and final layouts. Plates 3 and 4 show the final arrangement.

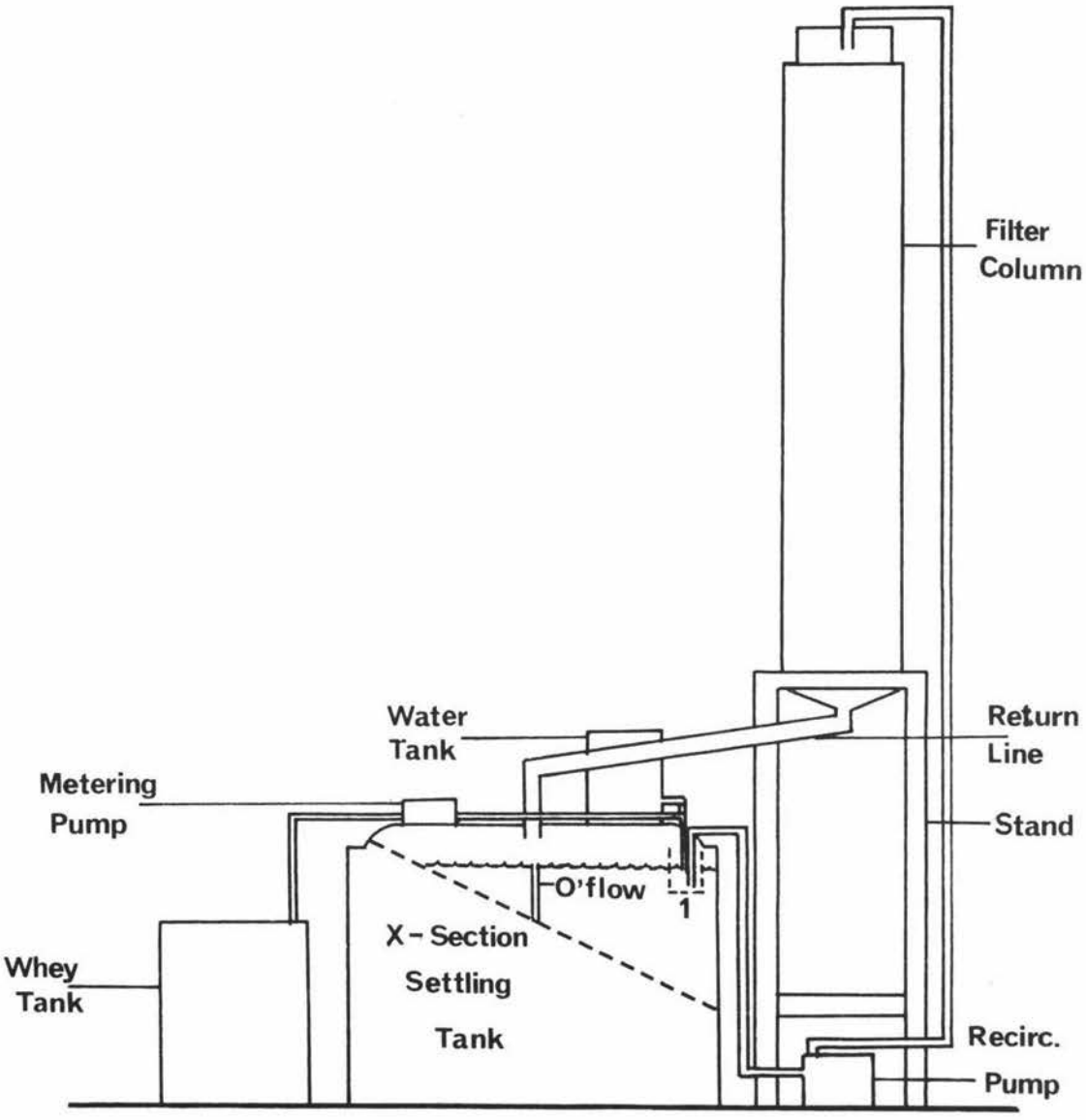
Details of the day-to-day operation of the plant and the methods of coping with minor problems are given in the Plant Log Summary in Appendix 2. Only the major changes are considered in the discussion below.

The plant was brought into operation by filling the settling tank to the 245 gallon mark with water incorporating 4.9 gallons of raw whey, and 2.5 litres of effluent from the laboratory plant as seed. This seeding was continued on a daily basis for two weeks. The development of a flora was noticed within two days of the start of operation. The feed to the plant was started on the fourth day, but proved too strong, and the whey rate was soon halved. In the first six weeks of operation an air pump was frequently used, as was the settling tank agitator, in an attempt to control the tendency for pH drop in this vessel. This settling tank arrangement proved unsatisfactory, especially when it was found difficult to remove excess sludge which floated to the top and created an odour nuisance.

Fig. 3

ORIGINAL PILOT

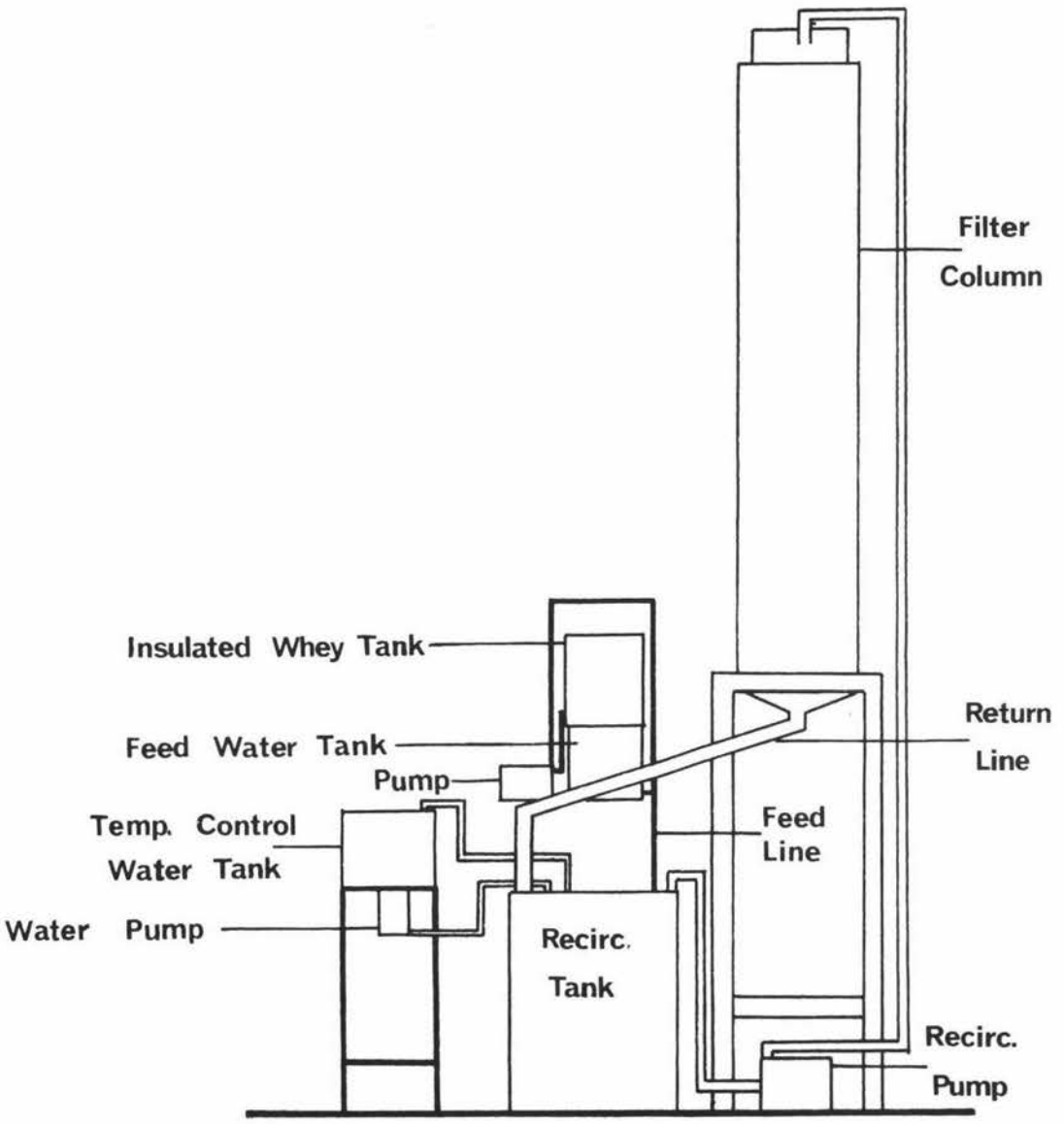
PLANT LAYOUT



Scale
1cm = 1′

1 Feed / Recirculation
Vessel

Fig. 4
FINAL PILOT
PLANT LAYOUT



Scale
1cm = 1′

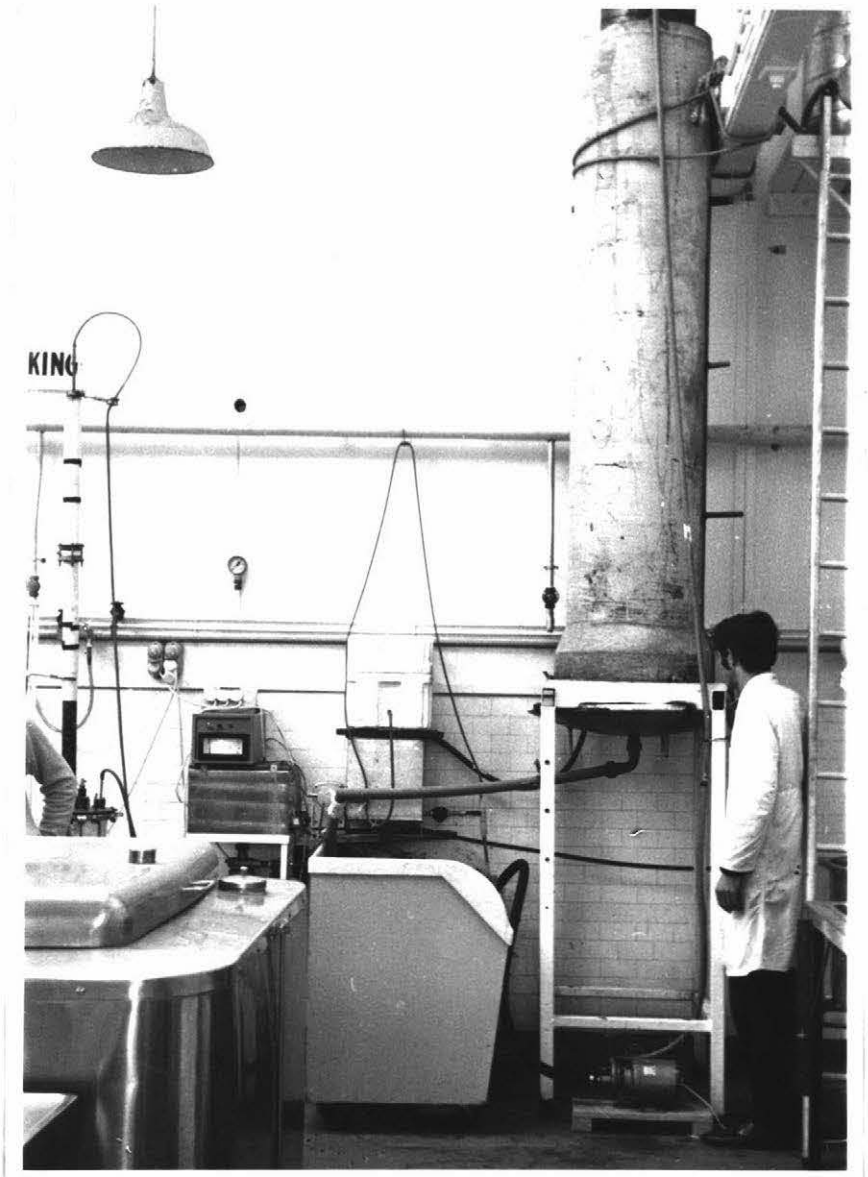


PLATE 3.

PILOT PLANT LAYOUT.

PLATE 4.

PILOT PLANT RECIRCULATION TANK.

In view of the paper by Harding (1952) the large settling tank was abandoned and a small (25 Igal) recirculation tank was fitted, with provision for settling the final effluent. This provision was used for only part of the operation as required. For test purposes the samples drawn could be settled in flasks as required. A new centrifugal pump was fitted to this arrangement; this pump had the flow range desired for later experiments. The plant performed much better under this flow pattern.

Another major modification to the plant was the conversion to a whey powder concentrate for feed. This became necessary when whey became unavailable at the end of the dairy season. The whey powder was reconstituted with water to form a concentrate 5.67 lb whey powder/10 gallons water i.e. 5.67% w/v. Details of whey and water flow rates are given in Table 10.

Other modifications include the fitting of a water-heated coil to the recirculation tank as a means of controlling temperature. The water temperature was controlled, and this gave a good control of tank temperature, as shown in Table 11, Chapter 2. A surface scraper device was fitted to the recirculation vessel to aid removal of washed-out growth from the column.

The outstanding difficulty in the operation of the column as an experimental unit was in the control of the recirculation pump flow rate at the value set by a gate valve in the outlet. Growth of flora occurred in the pump lines and also in the pump impellor itself, especially at the higher rates of flow. Cleaning the lines and the pump was found to be necessary approximately twice weekly. An attempt to overcome this problem was made by fitting a float constant level device to a feed tank at

the top of the column, the level controller switching the recirculation pump on and off. This was successful in controlling flow rate and cut back the cleaning required, but the pump motor had a mechanical start mechanism and the frequent switching caused heavy carbon build-up on the switch and frequent failure. Despite these difficulties, the plant supplied considerable experimental data.

As mentioned in the introduction, the investigation covered three major areas. The experiments of relevance to this chapter were the collection of data on the performance of the plant as a treatment of dairy waste. This is basically an assessment of the ability of the plant to remove BOD. BOD removal measurements were made at a variety of feed rates and strengths and of recirculation ratios. The experimental methods used are described in Appendix 1. The results are summarised in Table 10, and also in Table 6 and are presented in more detail in Table 11, Chapter 2.

Table 10.

Summary of Plant Performance.

PILOT PLANT

Whey Feed lg/day	Water Feed lg/day	Combined Feed Rate lg/yd^{3}day	Flow Rate incl. Recirc. lg/yd^{3}day	Recirc. Ratio	Average Feed BOD ppm	Organic Loading lb BOD/yd^{3}day	Organic Removal lb BOD/yd^{3}day	Efficiency %
3.51	81.81	162.8	3471	20.3:1	1620	2.637	1.686	64
3.51	81.81	162.8	4972	30.5:1	2078	3.381	2.734	81
3.22	85.83	169.9	5035	28.6:1	1172	1.991	1.574	79
3.22	85.83	169.9	6595	38.8:1	870	1.478	1.274	87
3.22	85.83	169.9	8915	51.5:1	1080	1.835	1.531	83
3.22	85.83	169.9	9420	54.5:1	1012	1.719	1.461	85

LABORATORY PLANT

cc/day	cc/day							
75	2425	170.8	58,880	344.5:1	612	1.04	1.01	97.3
75	2425	170.8	58,880	344.5:1	1237	2.13	2.10	98.8

SECTION D: THE PERFORMANCE OF THE EXPERIMENTAL UNITS IN THE TREATMENT OF DAIRY WASTES.

Both plants proved able to operate satisfactorily in terms of their ability to withstand varying feed strengths and, for the pilot plant, varying flow rates, and still produce effluents of reasonable quality.

The laboratory scale plant was very far from conventional design and it is unlikely that commercial plants would ever operate at the high flow rate of 59,000 Ig/yd^{3}day used, or as batch plants. Comparison of the performance with typical dairy waste treatment is therefore not particularly valid. However, it can be said that this plant consistently produced effluents of very high quality from compounded wastes of frequently high strength, and that the high flow rate used did not appear to be detrimental to the treatment. The plant had very few problems in its operation, and the use of lightweight foamed polystyrene packing media points to a possible commercial development.

The pilot scale plant was much closer to common design. The plant proved able to perform the type of treatment for which it was designed, in that it provided a 'roughing' type of treatment with a high removal of BOD on a weight basis per unit volume, but a lower efficiency in terms of percent applied BOD removed than many typical plants. The effect of the high flow rates used is discussed more fully in Chapter 2, but it can be said that at these flow rates no blockage of the column occurred, as would be expected with conventional plants operating at the high organic loadings used. The relatively large media and depth of column used probably also aided in this, the former by provid-

ing greater void space and the latter by increasing the vertical flow rate per unit of horizontal area and hence the flushing action. Comparison of the plant's performance with units described in the literature (Table 6) shows that, while the efficiency of the pilot plant was generally lower in terms of percent applied BOD removed, the removal of BOD on a weight basis per unit filter volume was generally higher. In conclusion, a plant designed on the basis of this pilot plant could be expected to give a satisfactory 'roughing' treatment to dairy factory waste.

CHAPTER 2.

PERFORMANCE-LOADING CHARACTERISTICS OF THE TRICKLING FILTER.

INTRODUCTION.

A major objective of the experimental work was to assess the effect of flow rate and recirculation-ratio variation on the performance of a dairy waste trickling filter operating at the high loadings used in the experimental units. It was hoped the data collected could be used for studying the many theoretical and empirical relationships available for performance-prediction, and testing their applicability to the experimental plants.

The majority of the operating equations used in filter design and performance prediction have been developed empirically from collected data and operating experience. Because of the absence of a theoretical basis for these predictions, and because of frequent contradictions among them, a great need exists for a unified, theoretically based approach to the analysis of trickling filter performance. There have been several attempts at this, and three of these are considered below. None has, as yet, had any wide practical application.

THEORETICAL APPROACHES.

Herbert (1961) and Herbert et al (1956) developed an approach for the analysis of continuous culture systems. This approach has been very successful and has had wide practical application. Its application to trickling filter analysis is not without problems. In classifying continuous culture systems,

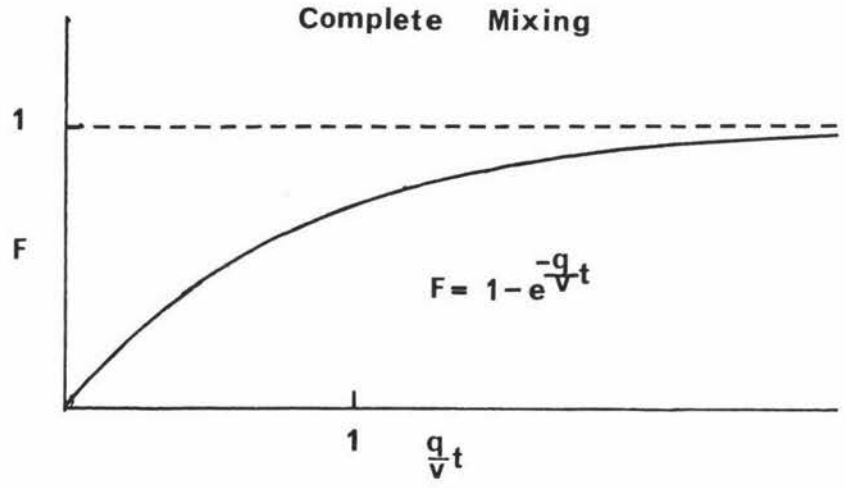
Herbert (1961) placed trickling filters in a grouping described as closed, heterogeneous and two-phased. This classification implies no cell loss from the system and an inability to reach steady state and true continuous operation. Both these implications are untrue of trickling filters, as also is the description as a two-phase system. Three phases, gas, liquid and solid, are important. Also, the Herbert-style analysis was originally developed for pure culture systems; trickling filters have a mixed culture.

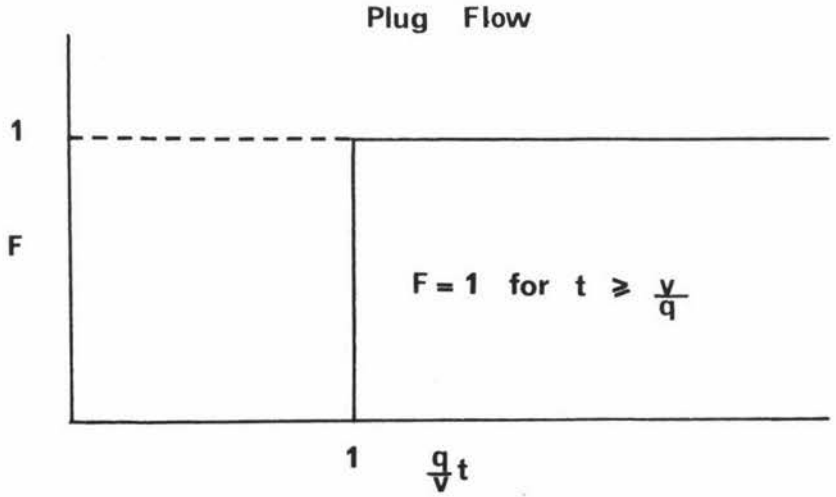
Despite these discrepancies, the basic style of approach used by Herbert can be applied to trickling filter analysis, as was demonstrated by Behn and Monodjemi (1968). The basis of the analysis is that the performance of a continuous culture system i.e. the removal or build-up of substrates or organisms depends on two main factors. These are the flow characteristics of the reactor and the kinetics of the reaction taking place.

The flow pattern can be expressed by the distribution of the residence times of the flow particles, according to Danckwerts (1953). The two extremes of flow type are plug flow and completely mixed flow, most reactors having a flow pattern between these. For completely mixed flow the age distribution of particles in the effluent ranges from zero to infinity, and may be represented by an equation and diagram as shown in Figure 5. For plug flow, all the particles have the same residence time, also shown in Fig. 5. If a first-order reaction occurs in the system, and P is used to denote the performance of the system i.e. the fraction of the original reactant concentration remaining in the effluent, then

$$P = \int_{t=0}^{\infty} e^{-kt} \frac{dF}{dt} \cdot dt$$

Fig. 5
PARTICLE RETENTION PATTERNS





F Fraction of Particles
v Reactor Volume
q Flow Rate
t Time

where k is the reaction rate constant and the other terms have meanings already defined. For complete mixing

$$P = \frac{1}{1 + \bar{t}\,k} \quad \text{where} \quad \bar{t} = \frac{V}{q}$$

For plug flow

$$P = e^{-k\bar{t}}$$

The usual technique for calculating the performance of a real system is, according to Behn and Monadjemi (1968), to base it entirely on plug flow or on completely mixed flow and to draw conclusions on this basis. A packed column such as a trickling filter generally approximates to plug flow, the approximation depending on the viscosity, density and velocity of the liquid and the size and shape of the packing.

In their approach to the reaction kinetics of a trickling filter system, Behn and Monadjemi used assumptions different to those of Herbert. This was probably necessary because Herbert's kinetic analysis was developed from pure culture, single substrate considerations. The starting points for the more complex situation of a trickling filter are that the kinetics of biological oxidation or organic material follow a first-order reaction, and that a direct consequence of the biological utilisation of organic material is the formation of new cells. Organic material concentration, as expressed by BOD, and cell concentration may be combined into one expression as done by Keshevan, Behn and Ames (1964)

$$\frac{dL}{dt} = -K\,L\,N$$

where L = BOD concentration
K = Rate constant
N = bacterial conc.

If a constant yield of bacteria for a given BOD reduction is assumed, then

$$\frac{dN}{dL} = - a \quad \text{where } a = \text{yield}$$

If the flow is assumed to be plug, then

$$\frac{dL}{dt} = - K L N$$

The integration of this expression is the major difficulty in the application of this type of analysis. Integration is simple if the reaction of BOD removal is simply first order. As already shown the efficiency may be expressed in this case by

$$p = \frac{L}{Lo} = e^{-kt}$$

P = fraction BOD remaining
Lo = initial BOD
L = final BOD

This is the starting point of analytical procedures used by Howland (1957), Schulze (1960) and Eckenfelder (1961). These are considered more fully below. If, however, an expression incorporating bacterial numbers as well as BOD removal, such as that of Keshevan et al above, is used, integration is much more difficult. The problem is to obtain meaningful values for N_o to N, for a system as complex as a trickling filter, where bacterial concentrations would be very difficult to determine. The Herbert style of approach, as developed by Behn and Monadjemi, is a useful one and will have more application as difficulties such as that above are overcome.

Fair, Geyer and Okun (1968) approached the whole issue of waste treatment kinetics in a new way, providing a possible basis for a unified approach to trickling filter analysis. The following discussion is based on this. The basic concepts of water and waste-water engineering can be included in a common mould, and because the principal operations and processes of treatment are slow, the rate at which they proceed, rather than the

position of final equilibrium is important. The time dependence of these rates falls into the area of reaction kinetics or transfer kinetics or both. An increase in speed is significant in that it allows a reduction in plant size.

Biological treatment can be considered as a series of interrelated operations, first in time and importance being the transfer of impurities from the wastes to the film. Second in time and, equal in importance, is the preservation of the quality of the contact between the wastes and the film, this being done by oxidation of the organic matter and by new cell synthesis. Interfacial transfer is considered to be the rate determinant, the purification pressure being a function of the concentration of removable substances and the removability of the fractions. A simple equation expressing BOD removal in a biological treatment plant is

$$\frac{dL}{dt} = k\,\phi(L)$$

where k = reaction rate constant

$\phi(L)$ = function of BOD conc.

t = treatment time

The following is presented by Fair et al as a general equation for waste treatment incorporating the relevant physical variables

$$\frac{dL}{dt} = k_o\left(\frac{A}{C}\right)\left(\exp\left[\frac{E}{R}\left\{\frac{T-T_o}{TT_o}\right\}\right]\right)\left(\phi[G]\right)\left(\frac{L_o-L}{L_o}\right)^n\left(L_o^L\right)$$

Where k_o = initial reaction rate prior to modification by physical effects

$\frac{A}{C}$ = interfacial contact area/volumetric capacity, and measures the effect of interface on k.

$\exp\left[\frac{E}{R}\ \frac{\{T-T_0\}}{TT_0}\right]$ gives the influence of temperature on k, according to van't Hoff - Arrhenius.

$\phi(G)$ gives the influence of conjunction opportunity on k. G is a velocity or shear gradient and depends on power input, viscosity and detention time.

$\left(\frac{L_0-L}{L_0}\right)^n \left(L_0-L\right)$ accounts for the variation in k with time or the relative amount of treatment accomplished, L/L_0

n = constant depending on reaction type.

As already mentioned, treatment is a series of interlocking, often concurrent operations and may be considered in steps or as a whole. Where a series of synchronous and interlocking reactions can be characterised by an overall reaction rate constant K the overall constant is the sum of the components

$$K = \Sigma k$$

On this basis, the general treatment equation above can be applied to trickling filter analysis by individually considering the component reaction and their contribution to an overall k, or their influence on k.

A general equation for trickling filter performance is a first order or retardent first order one of the type

$$\frac{dL}{dt_d} = K\left(\frac{L_0-L}{L_0}\right)^n \left(L_0-L\right)$$ where td = detention or contact time

for n = 1, integration yields $\frac{L}{L_0} = 1 - (1 + kt_d)^{-1}$

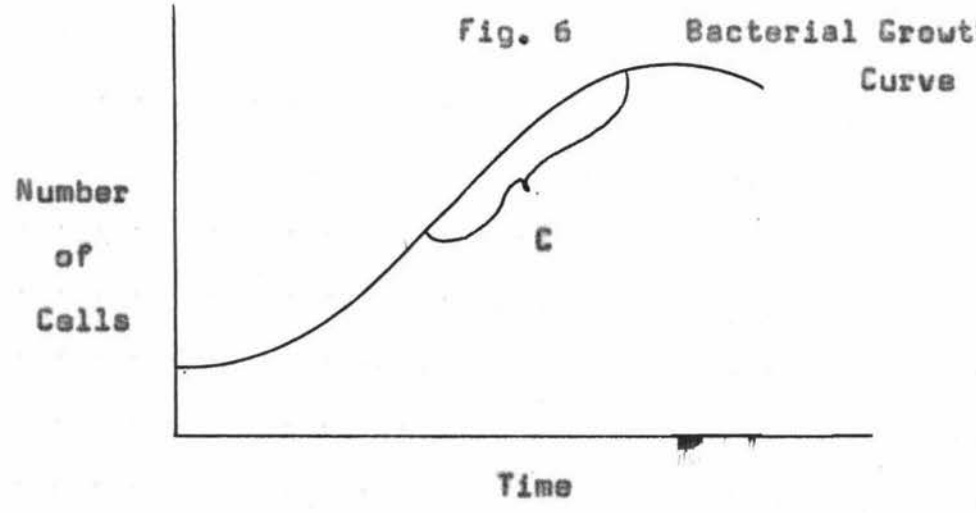
The following factors influence these basic equations as explained below:

Treatment Time. This depends on plant design and operation, and is best determined by tracer techniques, although formulae are available.

Temperature Effects. These may be considered to follow a van't Hoff-Arrhenius type relationship, which can be simplified to $\theta^{T_2-T_1}$ where $\theta = 1.040 \pm 0.005$ for θ^{T-20}, and T is in ^{o}C.

Nutrient Transfer and Substrate Utilisation. These may be incorporated on the basis that the controlling operations of biological treatment are (a) interfacial transfer and adsorption of organic wastes on biomasses (b) the support of these biomasses by nutrients. Interfacial transfer and adsorption can be expressed as a retardent first order reaction. So also can substrate utilisation by the biomass if it is assumed that the organisms are normally in section c of the growth curve in Figure 6. Hence the rate constants of these reactions can be directly incorporated into an overall rate constant, if they can be determined. Their determination is the main stumbling block in the application of this approach to trickling filtration, but is much easier with activated sludge, where bench-scale testing is relatively easy.

Fig. 6 Bacterial Growth Curve



Useful Power Dissipation. This may be expressed for filters as $k_g = \sqrt{gph/V_{td}}$

where p = proportion of useful power that enters the process

μ = absolute viscosity of the liquid

td = detention time

V = kinematic viscosity

h = depth

This factor k_g may also be incorporated in the overall constant.

Thus an overall purification equation with a theoretical basis for trickling filtration incorporating the effect of all the relevant variables is

$$\frac{dL}{dt_d} = \Sigma k \left(\frac{L_o - L}{L_o}\right)^n (L_o - L)\, \theta^{T-T20}$$

On integration this yeilds

$$\frac{L}{L_o} = 1 - \left(1 + n\theta^{T-T20} \Sigma k.t\right)^{-\frac{1}{n}} \quad \text{for } n > o$$

$$\frac{L}{L_o} = 1 - \exp - \theta^{T-T20} \Sigma k.t \quad \text{for } n = o$$

$$\frac{L}{L_o} = 1 - \left(1 + \theta^{T-T20} \Sigma k.t\right)^{-1} \quad \text{for } n = 1$$

The basic difficulty with this approach of Fair, Geyer and Okun is the determination of the numerical values for incorporation into the overall constant. This requires laboratory or larger scale testing for each situation which, as already mentioned, is more difficult for trickling filtration than activated sludge. Scale-up is a major problem, as detailed in Chapter 1. However, the unified approach to biological treatment is a very valuable one, highlighting the important physical variables, and is perhaps more adequate for the complex situation of a trickling filter than the more simple Herbert style of approach.

Atkinson and associates in a series of papers from 1962 - 1968 presented yet another approach to the analysis of trickling filters, on the basis of the development of a valid mathematical model. The need for a new approach to the problem was stated by Atkinson, Busch and Dawkins (1963). They claimed no existing theory or model adequately described the situation in trickling filtration. Swilley and Atkinson (1963) claimed all previous models for trickling filtration considered the reaction as being pseudo-homogeneous i.e. the biological oxidation process is implicity assumed to take place throughout the liquid film with no diffusional resistance to retard the rate of reaction. Oxygen availability was not considered limiting but food availability was the limiting factor. The model envisaged was a spongy, fibrillar biological mass through which the liquid floued. (Figure 7.) With this type of system recirculation can be shown to be advantageous, and is accountable by an increase in liquid film thickness creating an increase in effective reactor volume. If the reaction is not considered to occur throughout the film but rather at the interface between the microbial and liquid film i.e. is a surface reaction, the rate will be dependent on the rate of transfer from the liquid to the microbial layer and the rate of removal from this layer. (Figure 7) The effect of recirculation on performance for pseudo-homogeneous and surface models can be predicted as shown in Figure 8.

Atkinson, Swilley, Busch and Williams (1967) described experiments to determine which of the models was the valid one. The apparatus used was a sloping perspex plate on which was grown a thin layer of organisms; the medium used was glucose with nitrate as a source of oxygen. By varying the liquid flow rate and the

Fig. 7 REACTION MODELS

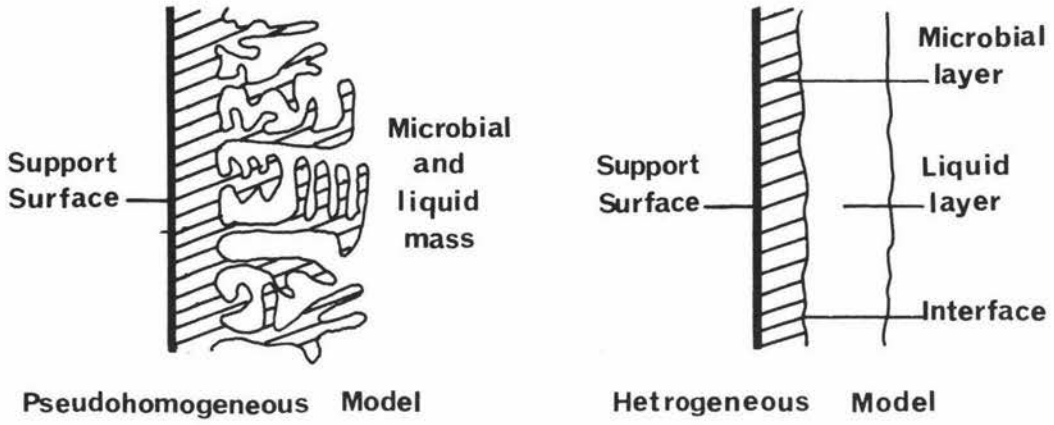
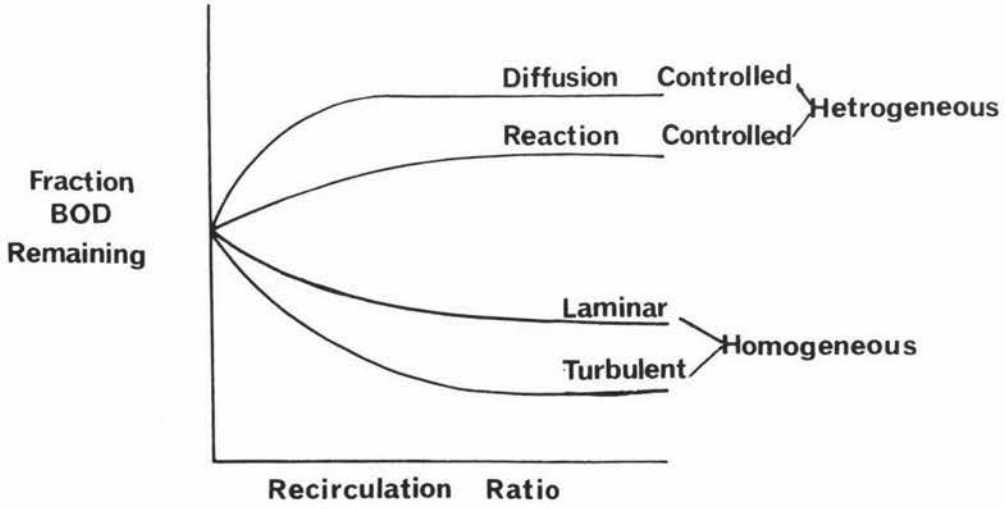


Fig. 8

RECIRCULATION AND PERFORMANCE



angle of the support surface the valid model was determined. The heterogeneous system with liquid phase diffusional resistance was decided on. It was concluded that liquid phase diffusion had considerable influence on reactor performance.

Atkinson and Daoud (1968) proposed as a model for the microbial mass individual organisms, with an outer inert diffuse zone and an inner reaction zone, scattered through a gel. They demonstrated that an analogy of a heterogeneous catalyst particle was valid, and that a diffusional limitation in the microbial mass was important.

Atkinson, Daoud and Williams (1968) presented a paper to co-ordinate the concepts of the previous papers, and the following complex equation was given for the prediction of reactor performance.

$$F = (1 + \tau)\, G(\eta, K, 0) - \tau$$

where F = fraction of reactant retained $\frac{L}{L_o}$

$\tau = \frac{a}{b\, L_o}$ where $r = a + b\, L^* = \frac{k_1\, d\, L^*}{1 + k_3\, L^*}$ = reation rate per unit area of biol film.

and L^* = conc. at interface

d = solid film thickness

$G = \tau + f$

where f = dimensionless concentration

$\eta = \frac{b\,\delta}{D}$ where D = diffusion coefficient, δ = liquid film thickness

$K = \frac{D\, z}{\delta^2} \cdot \frac{1}{Umax}$ where z = Reactor length

Umax = liquid film surface velocity

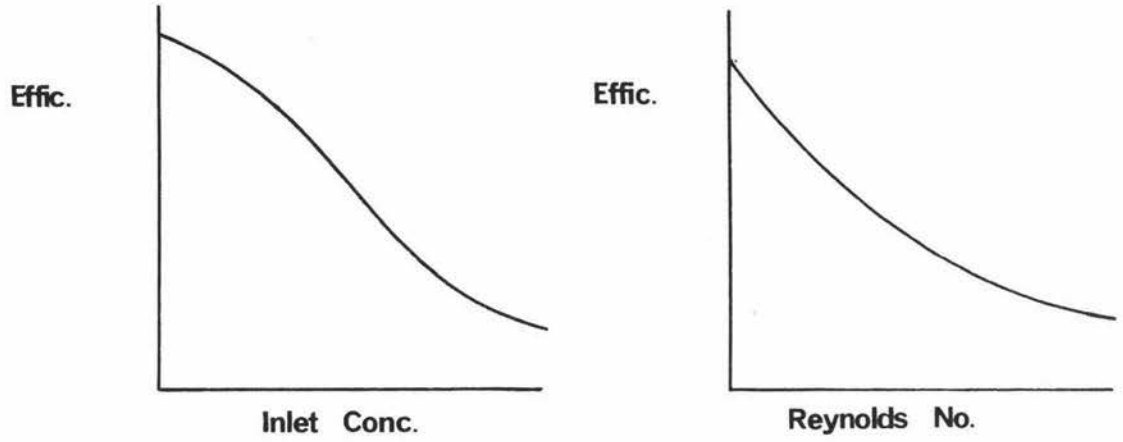
The effect of altering the variables of the equation was predicted by use of an iterative computer programme and the predictions

were compared with experimental results. In each case the predictions agreed with the experimental results, and the effects are presented diagramatically in Figure 9.

The authors concluded that they had developed a viable experimental tool for the study of microbiological reactions and a theory that allowed the biological coefficients to be determined and used in the design of biological film reactors. These conclusions would not appear to be entirely justified. Recirculation rate increase is considered detrimental to performance on the basis of the predictions of Atkinson et al, yet high recirculation is an established factor in the performance of high rate filters, permitting high loadings and high efficiencies. It contributes to uniformity in plant loading and flexibility of operation, and 'seeds' the raw waste. It also ensures that oxygen is not limiting. The conclusions of Atkinson et al are no doubt valid for their experimental equipment; the question is whether their model is a valid one for a trickling filter. On the basis of the information collected for the design of the plant used in this thesis, as given in Chapter 1, Atkinson's model would appear to be rather far removed from a true trickling filter. His treatment of the microbial mass and its likely effect on performance is not particularly adequate, and in some series of experiments the layer of organisms was deliberately scraped off to a thin film. No balanced flora of the type present in trickling filters could be expected to develop under such conditions. The theoretical approach of Atkinson et al would not appear to be the answer to trickling filter analysis, and it must be concluded that no unified theoretical approach of practical application to trickling filtration is yet available. Because of this, more

Fig 9

VARIABLE ALTERATION



If temperature increases, efficiency increases. If organism growth increases, efficiency initially increases, then becomes independent of growth.

conventional approaches with less theoretical backing are more common, and some of these are discussed below.

MORE CONVENTIONAL APPROACHES.

Behn (1963) and Behn and Monadjemi (1968) presented reviews of the operating equations that have been developed for trickling filters. The following discussion is based on these reviews, although other information is included.

The National Research Council (1946) presented one of the earliest operating equations. It was based on data collected from a number of U.S. military posts, and was given as

$$P = 100 / 1 + M i^{n}$$

where P is the percentage efficiency in terms of BOD removal
m and n are coefficients and i is the loading intensity.
For a single stage filter, the equation given was

$$P = 100 / 1 + 0.0085\sqrt{\frac{L}{VF}}$$ where L = applied BOD lb/ acre ft day
V = vol. filter acre ft.
F = recirc. factor.

Velz (1948) presented his well-known "basic law for the performance of biological filters" and stated that "the rate of extraction of organic material per unit interval of depth of a biological bed is proportional to the remaining concentration of organic matter, measured in terms of its removability" i.e.

$$\frac{dL}{dD} = - K L \quad \text{i.e.} \quad \frac{LD}{L_a} = \exp^{-KD}$$

where L_a = removable BOD applied K = rate constant
LD = removable BOD at depth D

Both Velz and Stack (1957) assumed there was a limiting load above which purification rate remained constant. Phelps and Velz (1948)

assumed this was a function of the storage capacity of the bed, and developed equations based on a first-order rate of removal of this. Stack considered the filter would saturate at various depths in sequence. Behn (1960) considered the whole theory of saturation required further verification.

Fair and Geyer (1955) first presented their theory of the rate of BOD removal being a modified first-order reaction. Both depth and contact time were considered possible alternatives for the basis of the rate expression, as the following equations show. As already shown, Fair, Geyer and Okun (1968) considered time alone.

$$\frac{dL}{dD} = - K L \left(\frac{L}{L_a}\right)^n \quad \text{or} \quad \frac{dL}{d_t} = - K L \left(\frac{L}{L_a}\right)^n$$

where t = contact time

Howland (1957), in contrast to Velz and Stack, supported time of contact rather than depth as the basic parameter, giving the following operating equation.

$$\frac{dL}{dt} = - K_t L \quad \text{i.e.} \quad \frac{L}{L_a} = e^{- K_t t}$$

Sinkoff, Porges and McDermott (1959) stated contact time in general could be given by

$$t = C_1 D \left(\frac{A_v}{Q}\right)^{n^1}$$

where C_1 is a constant

n^1 varies with filter medium

D = depth

Q = flow rate/area A_v = specific surface area

Eckenfelder (1961) used the above two equations in modified form, allowing for a decrease in active film with depth to obtain

$$\frac{L}{L_a} = e^{- K \frac{D^{1-m}}{Q^n}}$$

He further modified this to a retardent form to allow for different removal rates for different constituents to obtain

$$\frac{L}{L_a} = \frac{1}{1 + \frac{C_3 D^{1-m}}{Q n^1}}$$ where C_3, m, n^1 are constants

For rock filters treating domestic sewage he obtained

$$\frac{L}{L_a} = \frac{1}{1 + \frac{2.5 D^{0.67}}{Q^{0.5}}}$$ where D is in feet Q is in mgad

Ames, Behn and Collings (1962) developed a mathematical model for the trickling filter based on a chemical engineering type packed column, with a component in the liquor, BOD, being absorbed into the slime and undergoing a first order reaction. Through mass balances and partial differential equations they found

$$\frac{L}{L_a} = \beta + (1 - \beta) \exp - \frac{R}{R + 1} \cdot \frac{k . A_v . D}{\rho Q}$$

where $R = \frac{K}{\alpha k . A_v}$ and K = First order coefficient

k = mass transfer coefficient

α = equilibrium constant

β = non-transmissable BOD fraction

ρ = liquid density

Amado (1964) expanded the exponent as below

$$\frac{-R}{R+1} \quad \frac{k A_v}{\rho} = -\frac{1}{\rho} \Big/ \frac{\alpha}{K} + \frac{1}{k A_v}$$

This is in the form of an overall mass transfer coefficient

$$\frac{1}{K_m} = \frac{1}{k A_v} + \frac{\alpha}{K}$$

This overall coefficient consists of two individual resistances, on to adsorption $\frac{1}{k A_v}$ and one to reaction $\frac{\alpha}{K}$. Allowing for an increase in with depth and a decrease with flow rate he finally

obtained

$$\frac{L}{L_a} = \beta + (1 - \beta) \exp - K_1 \frac{D^m}{Q^n}$$ where K_1, m and n are constants.

For an eight foot experimental model, he evaluated the constants to obtain

$$\frac{L}{L_a} = 0.1 + 0.9 \exp - \frac{0.61\ D^{0.628}}{Q^{0.440}}$$

Galler and Gotass (1964) proposed the following operating equation, based on a large number of data and multiple regression analysis.

$$L = \frac{8.31\ L_a^{1.19}\ (1 + N)^{0.28}}{(1 + D)^{0.67}\ T^{0.15}\ [(1 + N)\ Q_2]^{0.06}}$$

Where L, L_a are in lb/acre day N = recirc. ratio

T in ^oC Q_2 = undil. hydraulic load, mgad

Recirculation

Few of the operating equations above specifically incorporate the effect of recirculation on filter performance. The first attempt to quantitatively define the effect of recirculation was given by the National Research Council (1946). They devised a recirculation factor which was the ratio by which the true volume of the filter could be multiplied to obtain the effective reactor volume. The number of effective passes through the filter was defined as

$$F^1 = 1 + \frac{R}{I}$$ where R = rate of recirc. flow

I = rate of influet flow

Hence $$F = (1 + R/I) \Big/ 1 + (1 - f) \frac{R}{I}^2$$

where f was the availability factor, accounting for the decrease in treatability with increasing recirculation. The maximum

recirculation ratio could thus be found from

$$\frac{R}{I} = (2f - 1) / (1 - f)$$

The equation of Eckenfelder (1961) can be modified by a recirculation factor $\frac{R}{I} = N$ to obtain

$$\frac{L}{L_a} = \frac{1}{(1 + N)\left(1 + \frac{2.5\ D^{0.67}}{Q^{0.5}}\right) - N}$$

where Q = diluted flow mgad

The equation of Amado (1964) may be similarly modified to obtain

$$\frac{L}{L_a} = \frac{0.1 + 0.9 \exp - \frac{0.61\ D^{0.628}}{(1 + N)\ Q_2}\ 0.44}{1 + 0.9\ N\left(1 - \exp - \frac{0.61\ D^{0.628}}{[(1 + N)\ Q_2]}\ 0.44\right)}$$

Where Q_2 = undiluted hydraulic load, mgad.

As has been shown, there are many operating equations available for the prediction of the performance of trickling filters. Their approach and basis is varied, as are their predictions. Empirically based equations will probably continue to be used until a satisfactory theoretical basis has been developed.

EXPERIMENTAL METHODS.

As indicated in the introduction, the objective of this section of the experimental work was to assess the effects of flow rate and recirculation ratio variation on the performance of the pilot filter. Basically, this was done by measuring BOD reduction (Appendix 1) at different rates of flow through the filter. By controlling the rate and strength of feed, temperature, and flow rate through the filter it was hoped the true effect of flow rate variation and recirculation ratio could be assessed.

Feed Rate and Strength

Whey or whey concentrate was fed at a constant rate by a metering pump and was mixed with a constant flow of water from an orifice in a constant head tank. Initially, raw whey of uncontrolled strength was used, but for most of the BOD testing a concentrated solution of 5.67% w/v whey powder in water was used. The concentrate was stored in a refrigerator until use, when it was placed in an insulated container with a twenty hour feeding capacity.

Flow Rate

This was controlled by a gate valve on the exit line from the centrifugal pump. The rate was set by measuring the rate of flow in the filter return, using the bucket and stopwatch technique.

Temperature

This was controlled by means of an immersed twenty-foot coil of ¾" copper pipe in the recirculation tank. Water from a

constant temperature tank was recirculated through this coil. The water temperature was generally 22 - 23°C, the aim being to control the waste liquor temperature at 20°C.

EXPERIMENTAL RESULTS.

These are summarised in Table 11. Control of some of the variables was not as successful as desired and this made interpretation of the results more difficult.

Temperature control was successful, as can be seen in Table 11. Control within the one to two degree variation shown was very reasonable for the system used.

The most difficult variable to control was the flow rate through the filter. As discussed more fully in Appendix 2, the growth of biomass in the recirculation pump and lines made it very difficult to control the output of the centrifugal pump for more than a few days. A constant displacement pump would have helped overcome this difficulty, but none was available in the desired flow range. Cleaning the pump and lines was necessary every four to five days, especially at the higher flow rates. Effective control was obtained at the highest flow rate by means of a constant head/float switch device. Despite the known variation of flow rates from their set points, the values for flow rate were taken as those set. This was justified as flow rate was checked every few days and the lines and pump cleaned as necessary.

BOD loading varied as shown in Table 11. The water and whey flow rates were found to be constant, and the variation must have been in the strength of the concentrate itself. The BOD test was not at fault, as duplicates were generally within

Table 11.

Plant Performance.

Date	Mean Daily pH	Mean Temp. °C	Flow Rate Ig/yd³ day	Feed Rate Ig/yd³ day	Recirc. Ratio Flow-Feed / Feed	Mean Feed BOD ppm	Mean Effl. BOD ppm	BOD Removed ppm	% BOD Removed	BOD Applied lb/yd³ day	BOD in Effluent lb/yd³ day	BOD Removed lb/yd³ day	Feed mgad (US)	Flow mgad (US)
16.3.	5.40	18	3471.2	162.8	20.32	1490	610	880	59.06	2.426				
17.3.	5.45	20	"	"	"	1470	630	840	57.14	2.393				
18.3.	6.30	19	"	"	"	1647	492	1150	69.82	2.681				
19.3.	6.00	21	"	"	"	1680	602	1078	64.17	2.735				
21.3.	5.70	21	"	"	"	1740	692	948	54.48	2.633				
26.3.	6.50	20.5	"	"	"	1660	475	1185	71.39	2.702				
31.3.	6.40	20	"	"	"	1520	570	950	62.50	2.475				
1.4.	6.40	21	"	"	"	1750	476	1274	72.80	2.849				
Mean		20				1620			63.92	2.637	0.9514	1.686	2.52	53.767
2.4.	6.50	18	4971.9	162.8	30.54	2041	447	1593	78.09	3.321				
3.4.	6.50	20.5	"	"	"	2390	550	1840	76.99	3.891				
5.4.	6.80	19	"	"	"	1800	425	1575	87.50	2.930				
Mean		19.2				2078			80.86	3.381	0.6471	2.734	2.52	77.013
9.4.	6.80	19	5034.6	169.9	28.63	1330	195	1215	91.35	2.260				
10.4.	6.75	18	"	"	"	1240	270	1070	86.29	2.107				
13.4.	6.90	19	"	"	"	1330	68	1262	94.89	2.260				
16.4.	6.90	21	"	"	"	880	146	634	72.04	1.495				
19.4.	6.90	21	"	"	"	1220	207	1013	83.03	2.073				
20.4.	7.10	20	"	"	"	1260	137	1123	89.13	2.141				
22.4.	6.80	21	"	"	"	1210	282	928	76.69	2.056				
23.4.	6.30	21	"	"	"	950	242	708	74.53	1.614				
24.4.	6.50	21	"	"	"	880	247	633	71.93	1.495				
25.4.	6.60	22	"	"	"	1170	305	865	73.93	1.988				
27.4.	6.20	21	"	"	"	1320	345	975	73.86	2.243				
29.4.	6.50	22	"	"	"	1220	197	1023	83.85	2.073				
30.4.	6.50	21	"	"	"	1220	257	963	78.93	2.073				
4.5.	6.00	21	"	"	"	1330	192	1138	85.56	2.260				
6.5.	6.10	20	"	"	"	1010	280	730	72.23	1.716				
9.5.	6.30	20	"	"	"	1170	380	790	67.52	1.988				
14.5.	6.20	20	"	"	"	1180	372	808	68.47	2.005				
Mean		20				1172			79.07	1.991	0.4167	1.574	2.63	77.983
19.5.	7.10	21	6595.4	169.9	38.82	980	87	893	91.12	1.665				
22.5.	7.05	20	"	"	"	760	132	628	82.63	1.291				
Mean		20.5				870			86.87	1.478	0.1941	1.274	2.63	102.160
29.5.	7.15	20	8914.9	169.9	51.47	1080	115	965	89.35	1.835				
1.6.	6.90	19	"	"	"	1080	237	843	77.55	1.835				
Mean		19.5				1080			83.45	1.835	0.3037	1.531	2.63	138.088
6.6.	6.90	20	9420	169.9	54.45	870	132	738	84.83	1.478				
8.6.	7.05	19	"	"	"	1030	52	978	94.95	1.750				
11.6.	7.10	20	"	"	"	925	137	788	85.19	1.572				
15.6.	7.15	20	"	"	"	1090	140	950	87.16	1.852				
21.6.	6.50	18	"	"	"	1490	405	1085	72.82	1.843				
26.6.	6.95	20	"	"	"	1070	162	908	84.86	1.818				
Mean		19.5				1012			84.97	1.719	0.2584	1.461	2.63	145.918
Grand		20.08												

five to eight percent, and the results in Table 11 are averages of these duplicates. Bacterial action in the concentrate probably caused the variation in BOD levels of the feed. Because of the simultaneous sampling of feed and effluent for BOD testing, it was still valid to calculate BOD reduction on the basis of these samples, despite the known day-to-day variation in feed BOD levels.

DISCUSSION OF RESULTS.

The results of the pilot plant experiments are plotted in Figure 10, to determine the relationship between BOD removal and recirculation ratio, the objective of this work. The scatter is considerable; this was expected in view of the difficulty in controlling the variables. Several attempts were made to fit a curve to the means at each recirculation ratio. By using the technique of curvilinear regression, as outlined in Appendix 3, a parabola with the characteristics given in Table 12 was obtained.

Table 12.

Equation of Parabola Through Means

$Y = 17.778 + 3.079\ X - 0.0342\ X^2$
where Y = % removal of applied BOD
X = recirculation ratio
Standard error of estimate (on means) = 2.1701%
Index of correlation = 0.99427

This fitted the means very well. Prediction of reactor performance on the basis of this curve would not give exact results, however, because of the great spread about the means. The

standard deviation of the widest spread is 9.33% BOD reduction about a mean of 79% i.e. at a recirculation ratio in the pilot filter of 28.6:1, only 68% of the predictions for BOD removal would be within 9.33% BOD reduction either way of the predicted mean using the equation in Table 12, assuming the distribution of BOD reduction to be normal. More data would have to be collected at different recirculation ratios, especially in the upper region, for the curve to be reinforced. However, despite the wide spread of results, the implications of the equation are worth noting. It predicts an optimum BOD reduction at a recirculation ratio of 45:1, with an equal rate of fall-off in BOD reduction about this optimum point. Recirculation ratios of up to 45:1 gave improved BOD reduction, but ratios higher than this had a detrimental effect on plant performance. This optimum recirculation ratio is much higher than used in normal practice. On the basis of this work it appears that recirculation ratios up to 45:1 would be advantageous in a dairy waste treatment system designed on a similar basis to the experimental pilot plant.

Predictions based on the theories and operating equations discussed in the opening sections of this chapter can be viewed in terms of the data collected from the pilot plant. For some of the theories and equations, the experimental data is not in a suitable form or is not complete enough and cannot be used. The Herbert style approach is ruled out for this as is the complex equation of Fair, Geyer and Okun. The data cannot be directly used in the equations of Atkinson et al, but the implication of these authors that recirculation is not advantageous in view of filter performance is not supported, although

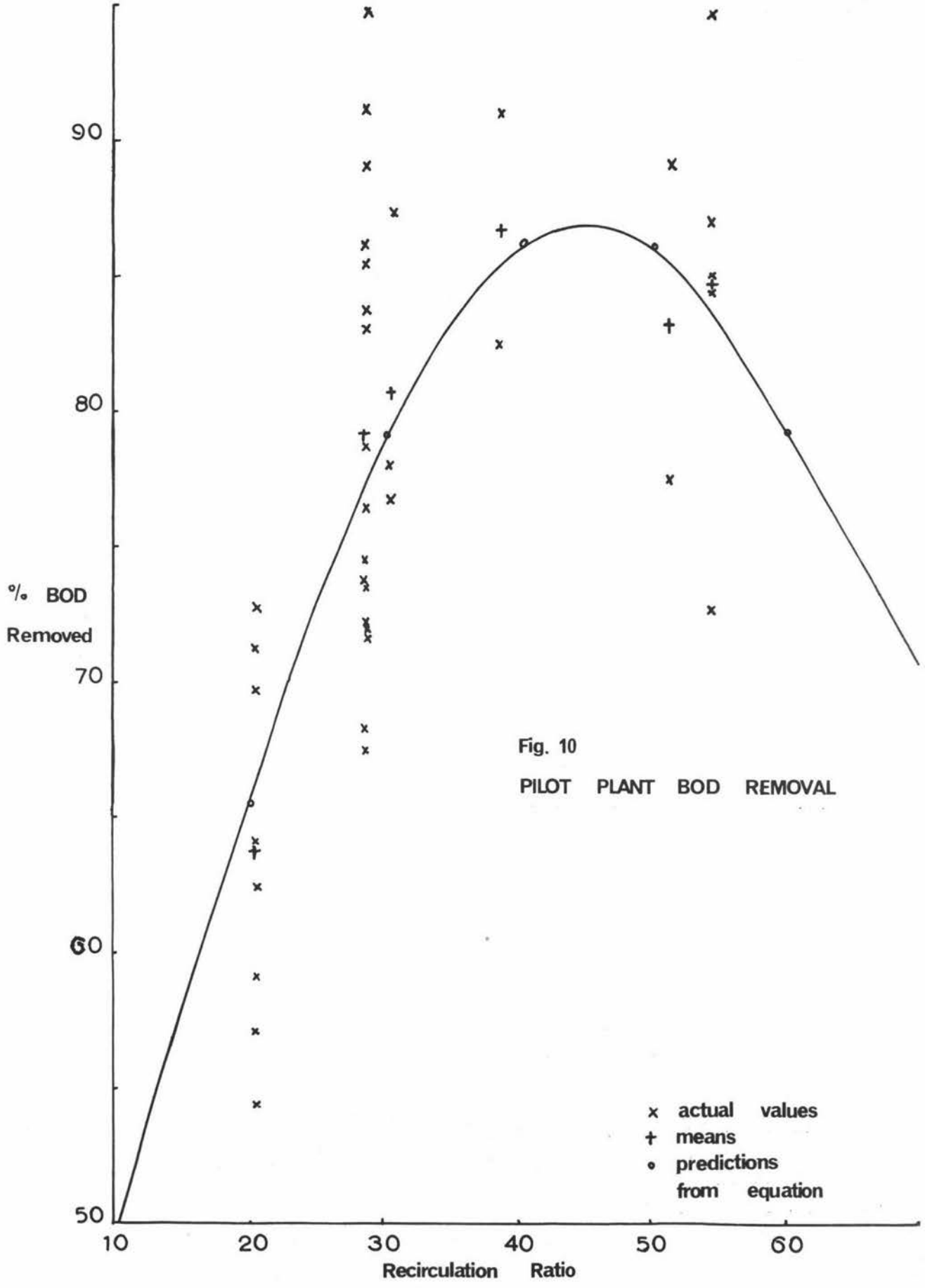


Fig. 10

PILOT PLANT BOD REMOVAL

The experimental data do predict an optimum recirculation ratio.

The data cannot be used in Velz's equation as depth dependence of BOD removal was not investigated. Similarly, because time of contact was not measured, the data could not be directly used in the equations of Houland, Fair and Geyer. However, on the assumption that time of contact is approximately inversely proportional to flow rate, the data do not support these equations, as attempts to plot flow rate versus efficiency on semi-logarithmic paper as a straight line were unsuccessful. As the equations are of exponential form, this plot should have been linear to support them.

The data were able to be used in the operating equations listed below. In many cases they required conversion into suitable units e.g. Ig/yd^{3}day to mgad (U.S.) The procedure in each case was to predict the efficiency in terms of BOD remaining in the effluent on the basis of the applied BOD and of the relevant parameters for that particular equation (Appendix 3) This prediction was plotted versus the major variable, flow rate. The <u>actual</u> mean BOD remaining was also plotted, and the graphs produced gave a comparison of the 'fit' of the prediction to the actual results obtained. The predicted points do not always lie on a smooth curve as would be expected. This is because the feed rate of flow varied slightly at certain total hydraulic rates of flow, and also because of the wide range in applied BOD loads at the different flow rates.

The following equations were plotted.

FIG. 11

NATIONAL RESEARCH COUNCIL (1946).

$$P = 100/\left\{1 + 0.0085\sqrt{\frac{L}{VF}}\right\}$$

where L = applied BOD lb/acre day
V = filter colume, acre feet
F = recirculation factor
P = % removal of applied BOD

FIG. 12

ECKENFELDER (1961), MODIFIED FOR RECIRCULATION.

$$\frac{L}{L_a} = \frac{1}{(1 + N)\left(1 + \frac{2.5\ D^{0.67}}{Q^{0.5}}\right) - N}$$

where D = depth, foot
N = $\frac{\text{Recirc. flow}}{\text{Infl.}}$
Q = dil. flow mgad
L = BOD remaining
L_a = BOD applied

FIG. 13

AMADA (1946).

$$\frac{L}{L_a} = \frac{0.1 + 0.9 \exp -\frac{0.61\ D^{0.628}}{(1 + N)\ Q_2}^{0.44}}{1.0 + 0.9\ N\left(1 - \exp -\frac{0.61\ D^{0.628}}{[(1 + N)\ Q_2]}^{0.44}\right)}$$

where Q_2 = undil. flow mgad

FIG. 14

GALLER AND GOTAAS (1964).

$$L = \frac{0.31\ L_a^{1.19}\ (1 + N)^{0.28}}{(1 + D)^{0.67}\ T^{0.15}\ (1 + N)\ Q_2^{0.06}}$$

where L, L_a are in lb/acre day
T in ^{o}C

Fig. 11
NRC EQUATION

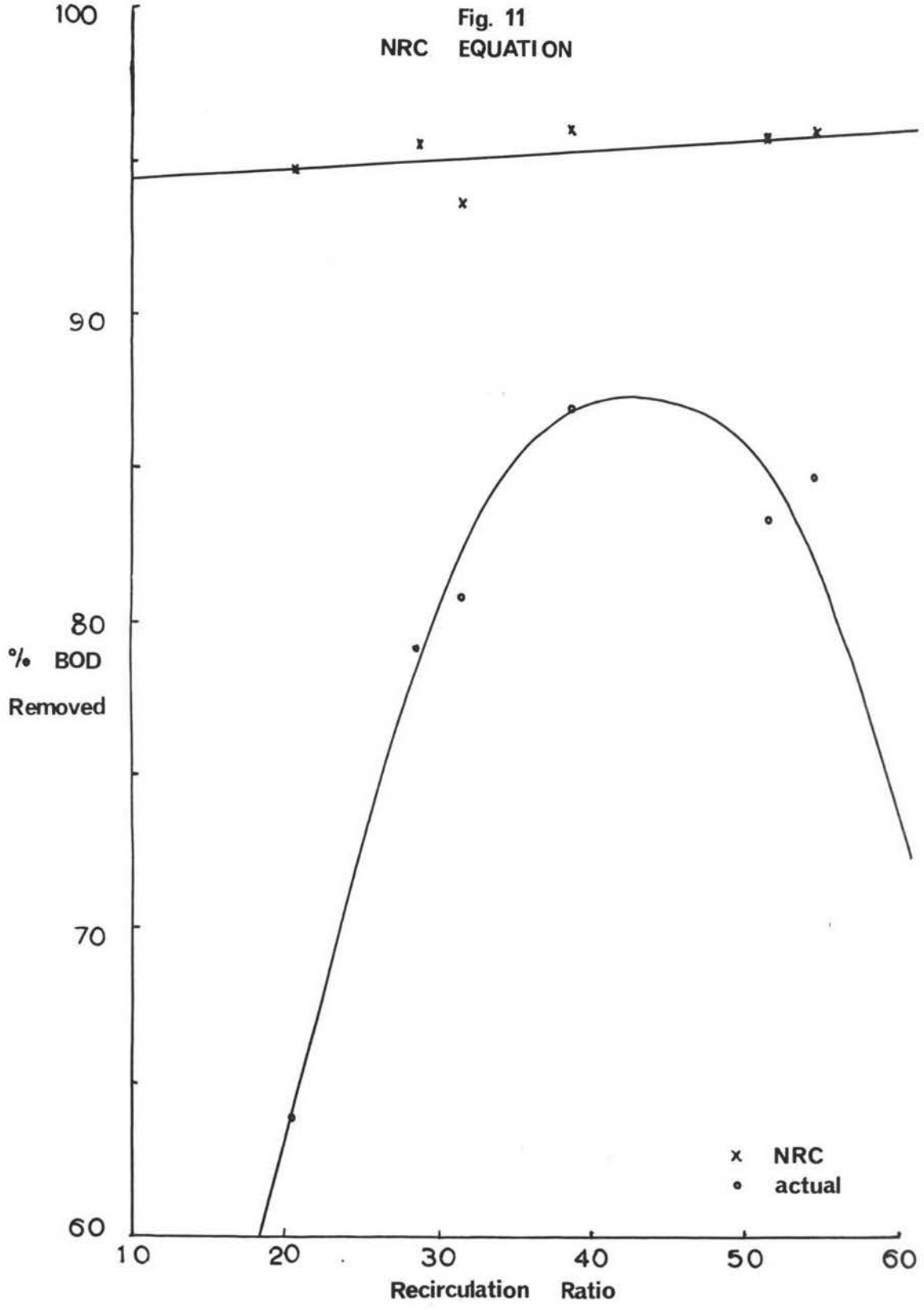


Fig. 12
ECKENFELDER EQUATION

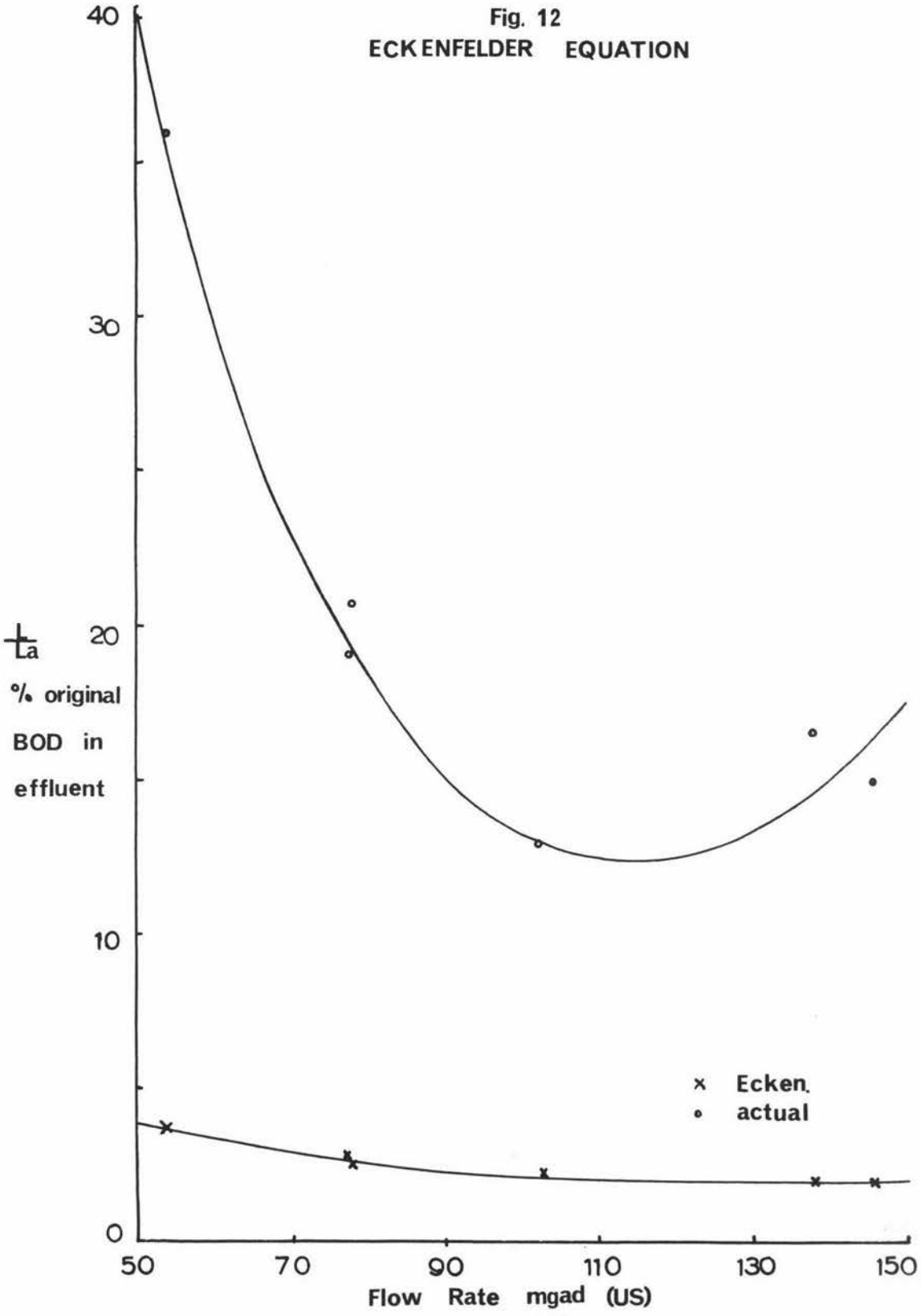


Fig. 13
AMADO EQUATION

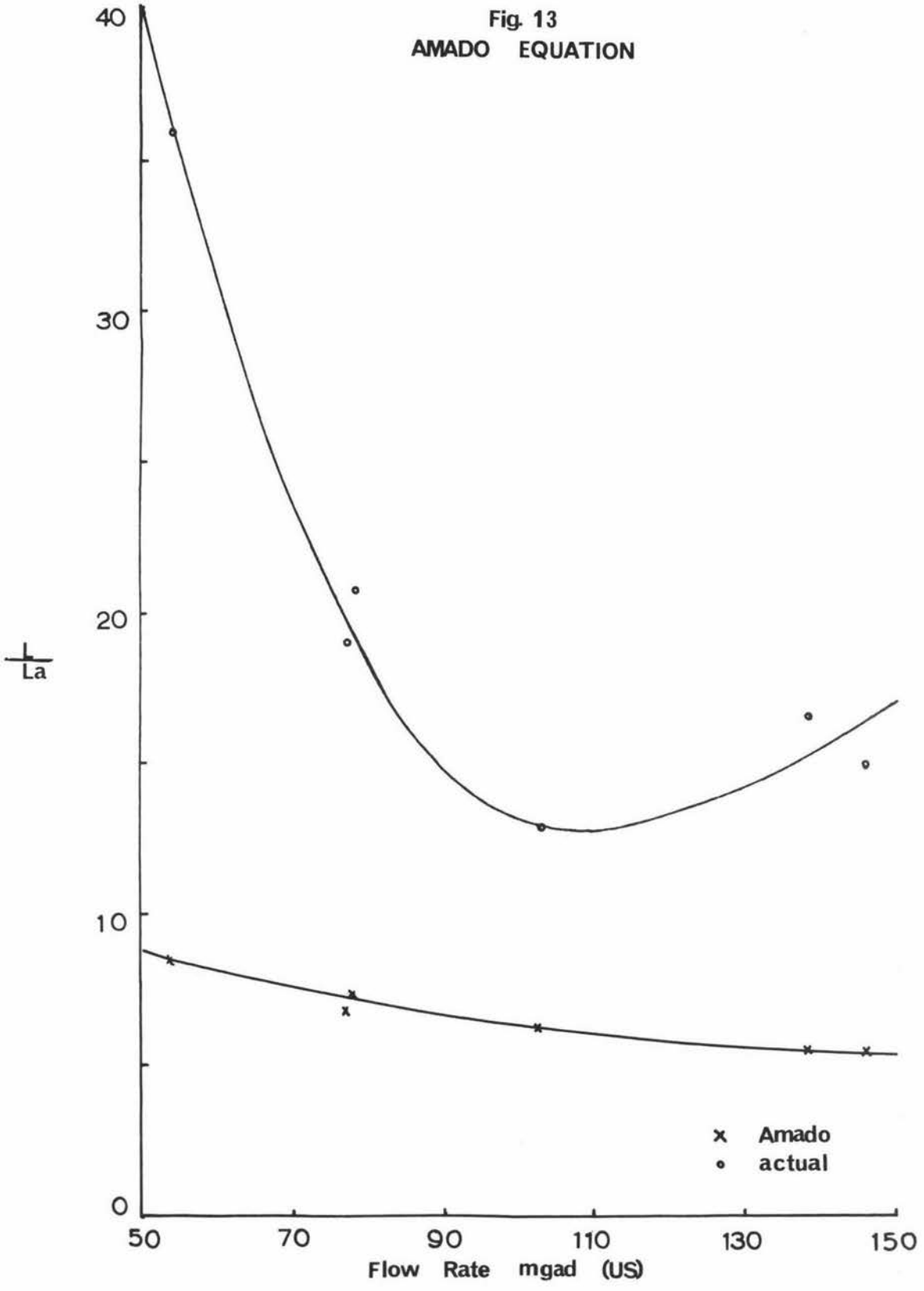
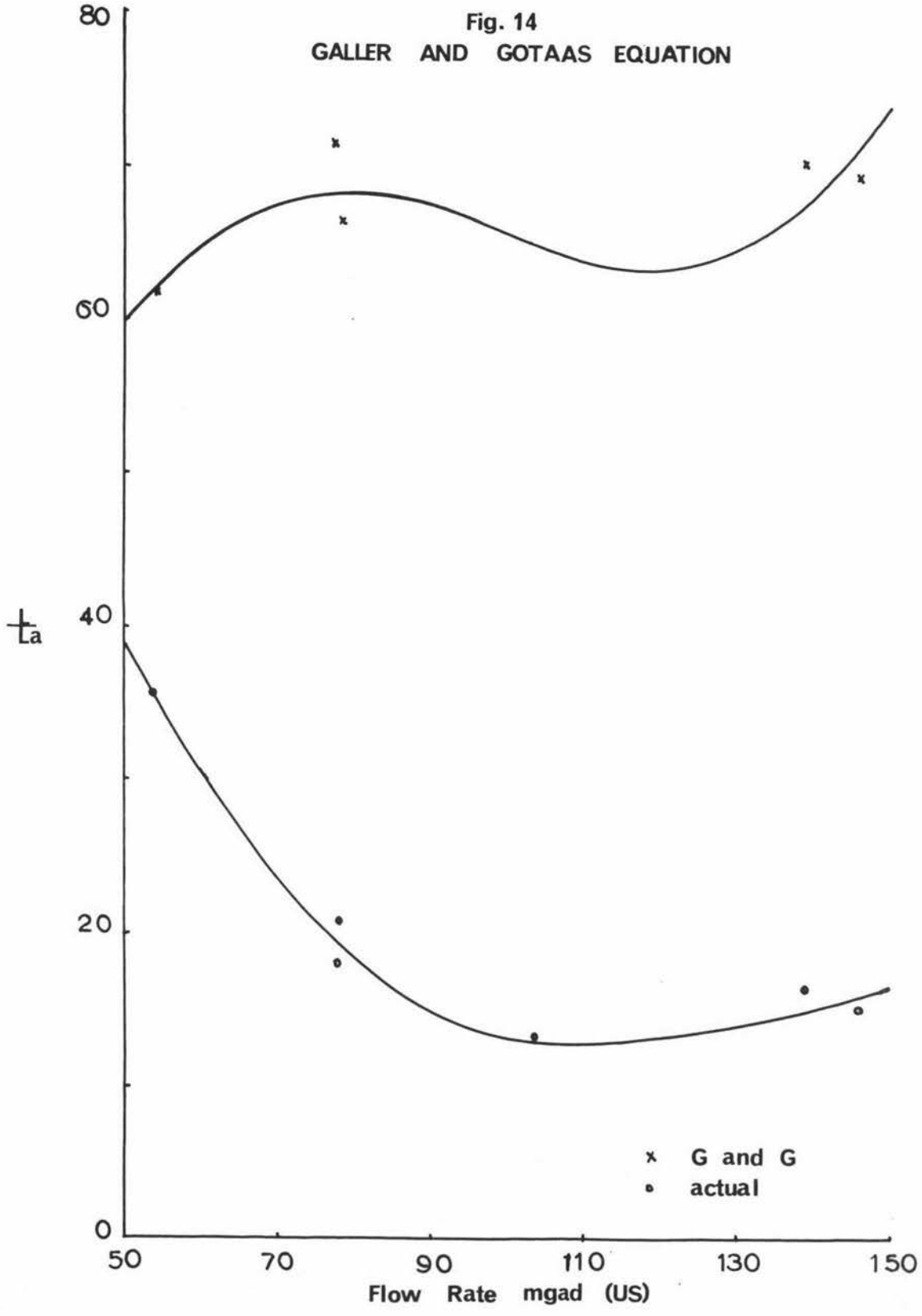


Fig. 14
GALLER AND GOTAAS EQUATION



None of the operating equations describes the actual situation at all successfully. The predictions also contradict each other. All but the Galler and Gotaas plot predict a plant performance better than that obtained. None predict the optimum flow rate found. All bar Galler and Gotaas predict an improved performance with increased flow rate.

It is not surprising that the experimental data and the predictions do not agree. Generally, the hydraulic and organic loadings used were above those for which the equations were developed. The waste being treated was different also, as the equations were usually for domestic sewage. The physical arrangement of the plant was unconventional for normal sewage treatment, in that settling was not provided for the recirculated flow. Feed strength varied considerably. The plant was under constantly varying conditions as the hydraulic loading was varied, and possibly did not always stabilise to the new conditions; certainly the biomass varied in composition, as discussed in Chapter 4. Disagreement between predicted and actual BOD removal efficiency could thus be expected.

CONCLUSIONS

The experimental data on BOD removal in relation to hydraulic loading rate and recirculation ratio do not support any of the theoretical and empirical predictions on filter performance in which they can be used. The data fit a curve described by the equation

$$Y = 17.778 + 3.079X - 0.0342X^2$$

where Y = % BOD removal

X = Recirculation ratio

This equation is specific to this filter and could not be validly applied to other plants. Recirculation ratios up to 45 : 1 appear to improve filter efficiency.

The data does not resolve the controversy over theoretical and empirical predictions for trickling filter performance. They do provide information on performance in loading regions higher than normally used for this type of filter. They serve to emphasise that it is extremely difficult to predict trickling filter performance, and that available operating equations should be applied with caution.

CHAPTER 3.

NITROGEN RELATIONSHIPS IN TRICKLING FILTRATION.

INTRODUCTION.

Nitrogen has been of interest in waste treatment since the very early stages of the development of waste treatment systems in the latter part of the nineteenth century. There are conflicting theories on the pathways followed by nitrogen in trickling filtration. Little work has been done on nitrogen relationships in dairy waste treatment by trickling filtration. The aim of this section of the work was to collect information on nitrogen relationships in dairy waste treatment by high rate trickling filtration, and also, to investigate the pathways followed by the nitrogen.

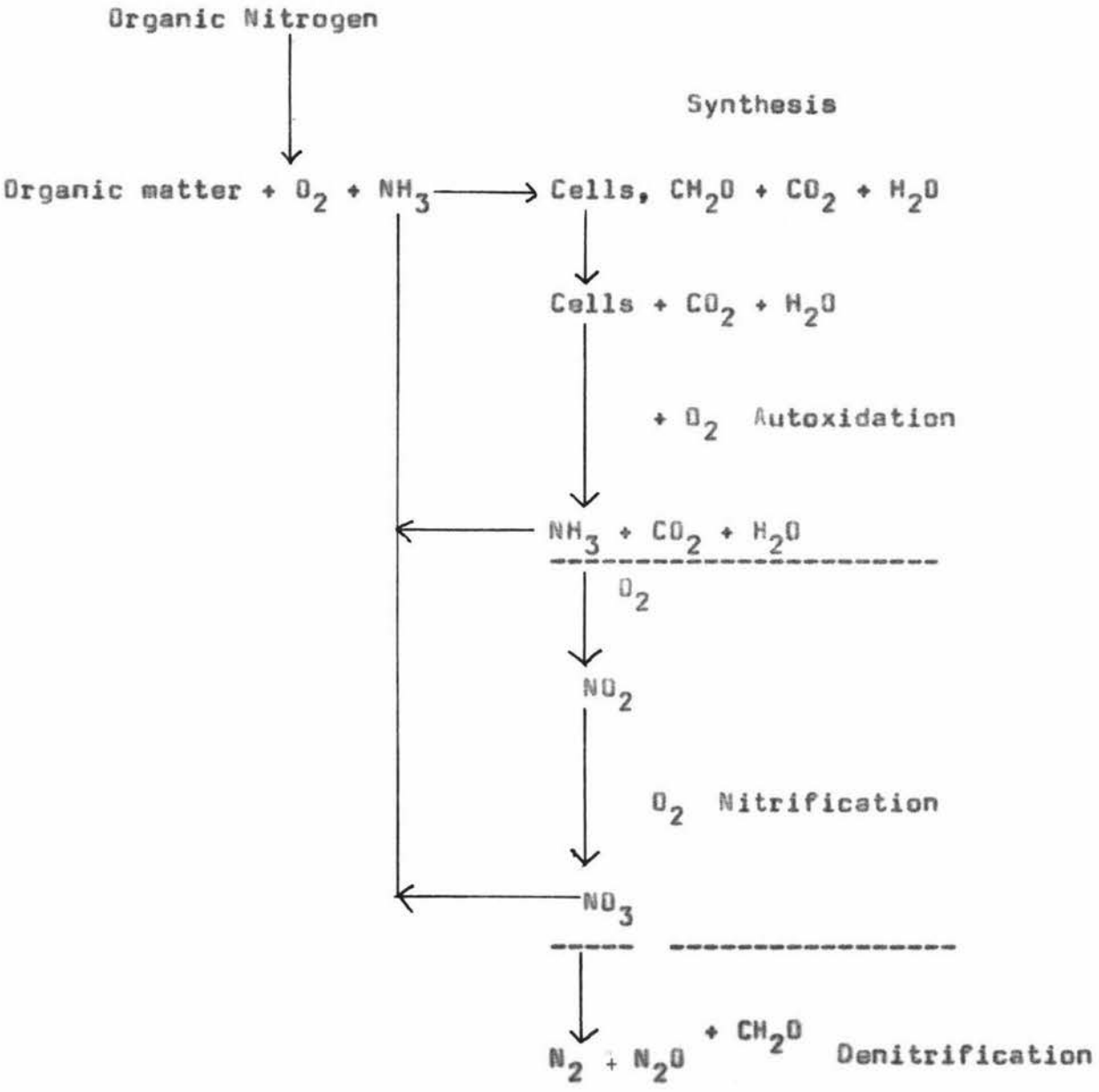
SURVEY OF PREVIOUS WORK ON NITROGEN.

The section of the nitrogen cycle of most importance in waste treatment is summarised in the Table below. This is based on work by Eckenfelder (1966).

As waste treatment methods have developed so has interest varied from part to part of this cycle. A characteristic of low rate filters is nitrification. Early work was concentrated on the nitrification step of the cycle. Interest in nitrification was particularly active because of concurrent interest in nitrification in soils. For the past thirty years, an increasing amount of attention has been given to nitrogen as a pollutant, and the whole of the cycle has become of greater importance, especially with regard to overall nitrogen removal.

Table 13.

Nitrogen Cycle.



Nitrification

Adeny (1890 - 1905) presented a series of papers on the biochemical oxidation of sewage and concluded that the oxidation

occurred in two stages. The first was oxidation of the carbonaceous material to carbon dioxide, ammonia and water. The second was oxidation of the nitrogenous material to nitrite and nitrate.

Beesley (1914) investigated the rate of nitrification of a number of substances and found they all nitrified at the same rate.

Barritt (1933) found the pH range for nitrification was 5.5 - 9.2, the optimum being 6.7 - 8.0. He claimed the inhibitory action of organic matter could be accounted for by accumulation of carbon dioxide and ammonia, and deficient aeration.

The concept that a trickling filter effluent should have a high nitrate level was discounted by Imhoff (1941). The proportion of nitrates to other nitrogenous materials was considered important, because the nitrates represent the end products of nitrogen breakdown. Therefore in a good quality effluent the proportion of nitrates should be high. More efficient use could be made of nitrogen if it was retained in the sludge for land application. Typically, an excess of nitrate production occurred when the filter was overloaded with sludge. This did not occur in high rate filters as excess sludge was washed out.

Viehl (1941), in a discussion of the presence of nitrates in effluents, concluded nitrate was not necessary, the true criteria of a good effluent being that it was clear, had a low BOD and some dissolved oxygen.

An investigation of factors affecting nitrification was made by Heukelekian (1942). He concluded that the presence of ample oxygen and ample numbers of nitrifying organisms was important. In contrast to earlier workers, he found nitrification was not adversely affected by carbonaceous material, provided sufficient nitrogenous material was present for its breakdown, and there was

adequate aeration. With low rate filters it was found that most of the organic material was removed by the upper section of the filter, nitrification occurring in the lower region, where oxygen was more available. As the load increased, the nitrification zone moved down the filter, until in high rate operation it was not present.

Tomlinson (1942) devised a method for measuring the nitrifying activity of a filter. This was by removal of baskets of medium from the filter and by incubation of suspensions of flora washed from these in aerated conical flasks, with 10 ppm nitrogen added as ammonium sulphate.

Heukelekian (1945) conducted a series of experiments for a comparison of low and high rate filters. He claimed the presence of nitrates in a filter effluent was not so much a measure of the purification as a criterion of the biochemical condition of the biomass. He confirmed that the production of nitrate and nitrite in high rate filters was uncommon, but was consistently high in low rate filters. A technique was devised for detection of the presence of nitrifying organisms. This consisted of making an aqueous suspension of sludge from the filters, incorporating some fresh sewage and aerating it for twenty-four hours and measuring the components. By this means it was confirmed that nitrifying organisms were well established in both types of filter. It was concluded that the differences between the two types were quantitative rather than qualitative and that in the high rate filter there was insufficient time for nitrate production.

Edmonson and Goodrich (1947) described experiments in a trickling filter specifically designed for nitrification. They found that after use as a nitrifying filter, a filter could handle

settled sewage at a much higher rate than a standard filter.

Mohlman (1948) further emphasised that little or no nitrification occurred in high rate filters.

The D.S.I.R. (U.K.) (1958) reported that nitrification in filters decreased markedly as temperature decreased.

A major paper on nitrification in the activated sludge process was presented by Downing, Painter and Knowles (1964). It included work on the growth of nitrifying bacteria in pure and mixed cultures, in a completely mixed aeration tank and in activated sludge plants. On the basis of this work it was concluded that

(a) A minimum period of aeration was necessary for nitrification. It had to be long enough for the nitrifying organisms to develop in the sludge.

(b) The minimum period was proportional to the BOD of the waste and inversely proportional to the sludge concentration.

(c) The growth rate constants of the nitrifying bacteria depended on the sewage and could not be predicted.

The Department of Civil Engineering, University of Newcastle-upon-Tyne (1965) presented a report on studies carried out on nitrogen in trickling filters. To determine the nitrifying capacity of the biomass, suspensions of the slime were aerated in the presence of ammonium sulphate. Denitrifying capacity was similarly estimated under anaerobic conditions. The concentration of nitrates in the effluent decreased as the hydraulic loading increased.

Balakrishnan and Eckenfelder (1969) (a) presented a paper on nitrification in trickling filters. They described work with a small scale filter $5\frac{1}{2}$" in diameter, 6' in length, packed with 1" Raschig rings and 1" Berl saddles. They developed a percent

nitrification relationship similar to that of Eckenfelder (1961) for BOD removal.

$$\% \text{ nitrification} = 1 - \frac{s}{s_o} = 1 - e^{-KD/Q^n}$$

where s = ammonia N influent conc.

s_o = effluent ammonia N conc.

K = reaction constant

D = depth

Q = hydraulic loading rate

n = constant

Their conclusions were

(a) That hydraulic loading rate had a profound effect on nitrification. (Note : Recirculation was not practised.)

(b) For the packing used with a specific surface of 67 ft^2/ft^3, at 30°C. n was 0.49 and K was 0.65

(c) There was very good correlation between specific surface of the media and nitrification rate constant.

(d) Temperature had a large influence on nitrification, the rate increasing with temperature rise. Depth of filter was also important, an increase improving the nitrification.

Nitrogen Requirements In Trickling Filtration

A definite level of nitrogen in relation to other components is required in a waste being treated by a trickling filter. There have been a number of reports in the literature as to the level of this requirement.

Levine (1929) found, in experiments on a wood lath type filter, that cheese factory wastes with a lower nitrogen content than normal dairy wastes required a greater filter depth for adequate treatment.

Barritt (1931) found wastes with a carbon/nitrogen ratio of above 15 - 20 : 1 contained insufficient nitrogen for oxidation of the carbonaceous material.

Weinberger (1949) found from activated sludge experiments that a BOD : Nitrogen ratio of 19.4 : 1 was necessary. This was reinforced by Kilgore (1953) who gave the ratio of 13.5 - 21.3 : 1 for high rate filters.

Symons and McKinney (1958) in a discussion of the biochemistry of nitrogen in the synthesis of activated sludge, commented that sufficient nitrogen must be available initially in this process for the synthesis demand to be satisfied. Inorganic chemical nitrogens were suitable for supplementation.

Johnston (1968) claimed that the maximum BOD : N ratio permissible in biological treatment was 17 - 20 : 1.

Goel and Gaudy (1969) used anhydrous ammonia as a nitrogen supplement for the treatment of carbohydrate wastes by activated sludge, and concluded the optimum nitrogen level depended on reactor detention time, allowable nitrogen leakage in the effluent and desired effluent quality.

Denitrification

Nitrogen is becoming increasingly important as a pollutant. Unoxidised nitrogen creates an oxygen demand on waterways, and both oxidised and unoxidised nitrogen contribute to the over-fertilisation or eutrophication of aquatic environments. This is thought to occur because nitrogen is frequently a limiting factor in the growth of algae and other aquatic plants in these environments. Nitrogen from treated and untreated wastes may overcome this limitation, and the decomposition of the consequent algal bloom

exerts too high an oxygen demand on the environment which may become anaerobic and "dead".

Nitrogen removal from the waste stream is therefore becoming more important, and may be accomplished in two main ways. The first is by conversion into settleable solids, i.e. by synthesis. The second is by denitrification of a nitrified effluent.

Denitrification was proven to be caused by micro-organisms by Hulme (1914). O'Shaughnessy and Hewitt (1935) found nitrogen was evolved when a nitrified solution was allowed to stand.

Ponninger (1943) stated that the fertilising effect of nitrogen on a stream was the same if the nitrogen was present as ammonia or nitrate, but the ammonia also removed oxygen. He claimed a well designed filter discharge filter sludge prior to any nitrification.

Delwiche (1956) said that in sewage treatment generally, provided oxygen was present, a vigorous nitrifying flora developed. The organisms were mostly obligate aerobes of the genera Nitrosomonas, Nitrococcus or Nitrobacter. Under anaerobic conditions, denitrification was caused by facultative anaerobic heterotrophs, which commonly produced nitrogen gas, utilising nitrate as an alternative hydrogen acceptor in the absence of oxygen.

Johnson and Schroepfer (1964) discussed the reduction of pollution and eutrophication by nitrification folled by denitrification, using a modified form of the activated sludge process.

Barth, Mulbarger, Salatto and Ettinger (1966) surveyed a number of municipal treatment plants for nitrification and denitrification. They found nitrogen removal to be erratic. Good nitrification was necessary for removal by denitrification. The best method

of denitrification was anaerobic treatment with a balanced source of oxygen demand material.

Eckenfelder (1967) stated the most promising method of nitrogen removal was by biological nitrification and denitrification.

Bringmann, Kuhn and Becker (1969) described a novel method for denitrification, by microbial synthesis of the nitrogen in the presence of an additional carbon source into biomass, which was suitable for animal feed.

Balakrishnan and Eckenfelder (1969) (b) conducted experiments on denitrification and found

(a) The rate increased as the concentration of the organic matter increased.

(b) Oxygen levels of 6 ppm prevented denitrification.

(c) The rate of denitrification was in direct proportion to the sludge concentration.

(d) By nitrification followed by denitrification 80 - 90% of the nitrogen could be removed.

Nitrogen Loss By Other Mechanisms

There have been a number of reports in the literature of a loss of nitrogen from trickling filters by some mechanisms other than synthesis and denitrification.

Rudolfs and Chamberlin (1931) reported on loss of ammonia nitrogen from an experimental trickling filter. The amount not accounted for by nitrification varied from 14 - 70%. Some of the ammonia was found in the air of the filter. Another possible loss was thought to be the formation of volatile ammonium compounds and oxides.

Zehender (1948) in a report on the elimination of nitrogen

and phosphorus in treatment plants, stated that the nitrogen removed decreased as load increased, and that not all the nitrogen was found in the sludge.

Beneden (1954) described experiments in which the nitrogen balance of aerated suspensions of sewage, enriched with nitrifying organisms and nitrogenous materials, was studied. The aim was to determine at what stage and by what means organic nitrogen was lost. He suggested the breakdown of amino acids by nitrifying bacteria in the presence of nitrates, accounted for the loss of nitrogen in gaseous form.

Leclerc (1959) stated that in a trickling filter there was a deficit in the nitrogen balance, mainly in the organic nitrogen. He claimed that, although nitrogen incorporation by the organisms of the filter was considerable, it did not account for the total loss of nitrogen. He discounted the hypothesis of ammonia volatilisation by proteolysis, as this would require a high pH, incompatible with normal filter operation. He proposed the following reaction between amino acids and nitrite to explain the nitrogen loss.

$$R\,NH_2 + H\,NO_2 = R\,OH + N_2 + H_2O$$

amino acid

This reaction was known, but acid conditions were necessary. Leclerc postulated the presence of a specific organism to cause this hydrolysis. If this reaction did occur, the role of nitrite and nitrifying organisms would be more important. The action of the nitrifiers would occur alongside that of the aerobic proteolytic organisms, producing nitrite which would be immediately utilised in the deamination reaction and would not be detected. As evidence for this hypothesis Leclerc stated

(a) The sudden cessation of the polluting powers of nitrogenous, undegraded material and the production of nitrogen gas on discharge to a waterway require explanation.

(b) The normal breakdown pathways of proteins and amino acids into amines, ammonia, acids, alcohols, alcohol acids and ethyl acids do not explain the nitrogen deficit. One possible mechanism is the complete oxidation by certain bacteria as shown by

$$4\ NH_3 + 3O_2 = 6\ H_2O + 2N_2$$

This requires a long time and the absence of nitrite and nitrate.

Summary of the Conclusions from the Survey of Previous Work

(a) Nitrification was important in low rate trickling filtration. The level of nitrate produced was frequently used as an indicator of purification and filter operation. A high level of nitrate is no longer considered desirable and it is unlikely that a high rate filter will produce a nitrified effluent.

(b) Certain minimum levels of nitrogen are required in relation to the organic load for adequate treatment. A typical maximum ratio of BOD : N is 20 : 1.

(c) Nitrogen loss from trickling filters and their effluents is becoming increasingly important. It occurs by denitrification of the effluent, synthesis of biomass and by other mechanisms.

EXPERIMENTAL.

The experimental work was done in two major sections, and description of the techniques used, results obtained and diseussion of these results is divided into these two sections. The first describes the investigation of the nitrogen relationships in the pilot plant at different recirculation ratios. The second describes the study of overall nitrogen balance by the use of suspensions of flora from the pilot plant.

SECTION ONE.

NITROGEN RELATIONSHIPS IN THE PILOT PLANT.

The technique used here was to simultaneously collect samples of pilot plant feed, effluent from the filter column and effluent from the recirculation tank. The samples of 100 ml were preserved with approximately 1500 ppm of concentrated (36 N) sulphuric acid and stored at 2°C. until analysis for total Kjeldahl nitrogen, ammonia cal nitrogen, nitrite nitrogen and nitrate nitrogen, as outlined in Appendix 1. Initially, the samples were analysed for Kjeldahl nitrogen on the day of collection. It was found that this was giving very variable levels for the column and plant effluent samples. This was probably caused by unsettled suspended material. For the latter part of the experiments, overnight settling was allowed prior to any analysis. Generally, a total of twelve samples could be analysed in one day. The maximum storage time for the preserved samples was twenty-four hours (Appendix 1). With the overnight holding found necessary for the Kjeldahl nitrogen, the number of samples in any one batch was limited to twelve. Centrifuging the samples prior to Kjeldahl analysis

would have overcome the difficulty with suspended material, but this would have given an artificially low suspended matter level and it would not have been possible to obtain a meaningful indication of the nitrogen removal being accomplished.

Results and Discussion

The results are summarised in Table 14. Presentation as mean and range at each recirculation ratio for the feed, column effluent and plant effluent indicates the typical nitrogen level and the spread of results about this. The percentage removal indicates how the level of that particular form of nitrogen varies from the plant feed to the plant effluent. It was calculated on the basis of the difference between levels in the effluent and influent expressed as a percentage of the influent. The manner in which each form of nitrogen varied with flow rate is considered in turn below.

Total Kjeldahl Nitrogen. This represents the nitrogen present as organic nitrogen and as ammoniacal nitrogen. Hence, by substraction of the ammoniacal nitrogen separately determined, the organic nitrogen could be determined. This was not normally worthwhile, as the organic nitrogen was generally much higher than the ammoniacal nitrogen, and the ammoniacal nitrogen levels did not fluctuate widely.

Consideration of the feed Kjeldahl nitrogen levels and the BOD : N ratio reveals that the plant was consistently operating at ratios above the maximum permissible value found in the literature of 20 :1. Despite this, the plant was able to consistently produce a satisfactory level of BOD removal, as discussed in Chapter 1. Also, Kjeldahl nitrogen was consistently present in both the column

Table 14.

Pilot Plant Nitrogen Levels

Flow Rate lg/yd^{3}day	Recirc. Ratio	Sample Type	No. of Samples	Total Kjeldahl Nitrogen, NH_3 + Org. ppm Mean	ppm Range	Nitrate Nitrogen ppm Mean	ppm Range	Nitrite Nitrogen ppm Mean	ppm Range	Ammoniacal Nitrogen ppm Mean	ppm Range
5034.6	28.63	PPF	11 [1]	55.1	38.0	0.25	0.35	0.01		4.1	4.5
		PPE	11 [1]	32.2	67.0	0.03	0.10	0.02		2.9	6.0
		PPC	11 [1]	27.9	66.0	0.10	0.30	0.06		2.3	6.4
% Reduction				41.5		88.0		100 (increase)		35.6	
$BOD_F : N_F$				21.3							
6595.4	38.82	PPF	12 [1]	34.5	30.9	0.22	0.35	0.005	0.03	4.6	8.7
		PPE	12 [1]	20.3	19.0	0.07	0.10	0.005	0.06	1.0	2.0
		PPC	12 [1]	20.0	20.0	0.06	0.20	0.03	0.07	1.2	4.7
% Reduction				41.1		68.2		Nil		78.3	
$BOD_F : N_F$				25.2							
8914.9	51.47	PPF	21 [2]	40.5	28.0	0.21	0.35	0.005	0.04	1.7	2.1
		PPE	21 [2]	10.4	19.0	0.08	0.25	0.001	0.09	0.8	1.6
		PPC	21 [2]	12.0	22.0	0.07	0.10	0.03	0.11	0.6	2.0
% Reduction				74.3		62.0		80.0		53.0	
$BOD_F : N_F$				26.7							
9420.0	54.45	PPF	6 [3]	39.7	24.0	0.27	0.05	0.00	0.01	2.1	1.1
		PPE	6 [3]	4.4	5.0	0.10	0.10	0.01	0.03	0.7	0.7
		PPC	6 [3]	7.6	4.5	0.10	0.10	0.01	0.4	1.0	1.4
% Reduction				88.9		63.0		Increase		66.7	
$BOD_F : N_F$				25.5							

NOTE: PPF means Pilot Plant feed sample
PPE means Pilot Plant effluent sample
PPC means Pilot Plant column effluent sample
$BOD_F : N_F$ means feed BOD : feed Nitrogen ratio

1 These samples were not settled overnight prior to analysis
2 Sixteen of these were settled overnight and only these results are included in the Kjeldahl nitrogen figures.
3 These samples were settled overnight.

and plant effluents. This would suggest that the ratios specified in the literature did not apply to the pilot plant under its operating conditions, as the presence of this form of nitrogen in the effluent would indicate not all the feed nitrogen was required for BOD removal.

The effect of recirculation ratio on the removal of Kjeldahl nitrogen was not clearly indicated on the basis of the results obtained. This was because overnight settling of the samples was practised at only two of the recirculation ratios employed. The results obtained prior to this cannot be validly compared with these, as the presence of unsettled material gave low percentage removal. On the basis of the results from when settling overnight was practised, it would appear that the higher recirculation ratio and flow rate gave a higher percentage removal of Kjeldahl nitrogen. Also on the basis of these results, removals of at least 70% total Kjeldahl nitrogen were likely in the pilot plant.

Ammoniacal Nitrogen. These results, and also the nitrite and nitrate results, were considered to be unaffected by whether or not the sample was settled, as the nitrogen would be in solution in these forms. Also, centrifuging of the samples prior to analysis for ammonia, nitrite and nitrate was frequently practised to avoid cloudy interference in the analytical methods used.

Ammoniacal nitrogen levels in the feed were consistently higher in the feed than in the effluents, the levels in the column and plant effluents generally being similar. Because there was no evidence of nitrification, the bulk of the ammoniacal nitrogen was probably being utilised in the synthesis of biomass on the column.

Recirculation ratio and nitrogen level did not appear to be related.

Nitrite Nitrogen. The levels in both the feed and the effluents were generally very low. They were consistently higher in the column effluent than the plant effluent, and the levels in both were sometimes higher than in the feed. This would suggest a very slight formation of nitrite in the column, but this cannot be definitely stated because of the very low levels that were being measured. Again, recirculation ratio and nitrite levels were not apparently related.

Nitrate Nitrogen. As with nitrite nitrogen, the levels were generally very low. The levels in the feed were consistently higher than in the effluents. The nitrate in the feed was probably from the dilution water from the header tank, as this was found to have nitrate levels of the same order. The levels in the column and plant effluents differed very slightly. Higher removal of nitrate was accomplished at the lower recirculation ratios, but in terms of actual quantities removed, the differences were very minor. There was no evidence at any stage of nitrate production and nitrification. This was expected on the basis of the literature which predicted that nitrification was likely only in low rate filters.

Conclusions

(1) The pilot plant operated consistently at BOD : N ratios in the plant feed higher than those recommended in the literature, but despite this, dissolved organic nitrogen and ammoniacal nitrogen were always present in the effluent. The BOD removal in the plant was satisfactory at this high ratio. It

would appear that the recommended BOD : N ratios did not apply to the pilot plant.

(2) Typically 70% of the incoming nitrogen was removed by the filter.

(3) There was no evidence of nitrification in the pilot plant.

(4) Flow rate and recirculation ratio had no apparent influence on ammoniacal, nitrite and nitrate nitrogen. All of these generally decreased in level on passage through the plant. The influence of recirculation ratio on organic nitrogen levels was unclear.

SECTION TWO.

STUDY OF THE NITROGEN BALANCE.

It was impossible to study the nitrogen balance on the pilot plant itself, as it was not possible to determine the quantity of nitrogen being converted into bulk biomass. The technique used was the incubation of suspensions of flora from the column in shake flasks, the nitrogen concentrations in samples from these being measured. The experimental procedure is outlined below:-

(1) Two gallons of liquor from the recirculation tank were settled for ten minutes and the supernatant siphoned off.

(2) The settled material was centrifuged at 3500 rpm for three minutes and the sludge recovered.

(3) Twenty grams of sludge were weighed into each of two 500 ml conical flasks.

(4) The sludge was diluted to 500 ml, with tap water for the control flask and with a solution of the additive for the experimental flask.

(5) The flasks were placed in an orbital incubator at 20°C. and 250 rpm for the aerated experiments, or in a standard incubator at 20°C. for the non-aerated experiments.

(6) Samples of 100 ml were taken initially and approximately every hour for the duration of the experiment, preserved with sulphuric acid and stored in the refrigerator for analysis the following day.

In contrasts to the determination of total Kjeldahl nitrogen on the pilot plant samples, the shake flask samples were well shaken prior to measuring the volume for total Kjeldahl determination. This was so any change in the total nitrogen balance could be determined. Where dissolved nitrogen was determined, the suspended matter was first removed by centrifugation. As with the pilot plant experiments, the maximum number of samples that could be analysed in a batch was twelve, and generally only twelve samples were drawn for each experiment with its particular additive. The conditions and additives used are shown in Table 15.

Results

The results are summarised in Table 16. The procedure used in handling each batch of results is shown below. Basically an attempt was made to develop a correlation between the particular nitrogen level and time of incubation. This was to determine the alterations in the levels of the various forms of the nitrogen.

Firstly, an F test at the 0.05 level of significance was used to determine if there was any validity statistically for distinguishing between the results for the control samples and the experimental samples i.e. those with an additive.

Table 15.

Compounds Added.

Aerobic (Aerated)	Anaerobic (Non-aerated)
Ammonium chloride and sodium nitrite	Ammonium chloride and sodium nitrite
Ammonium chloride and sodium nitrate	Ammonium chloride and sodium nitrate
Sodium nitrite and nitrate	Sodium nitrite and nitrate
Sodium nitrite	
Sodium nitrate	
Ammonium Chloride	
Casein	
Lactalbumin	
Lactose	

Secondly, the hypothesis that the regression coefficients for the data were actually zero i.e. no correlation existed was tested, again using an F test at the 0.05 level. The data were combined for this test if no valid difference had been found in the first test.

For the results where a valid regression existed, the equation of the linear regression line was calculated and the line drawn. These are shown in Figures 15 - 20. The equations of the lines are given in Table 16. If the regression was not valid, the results are expressed as means and ranges or as means, ranges and standard deviations in Table 16. A more simple presentation of results is given in Table 17.

Table 16.

Nitrogen Levels in Shake Flasks.

Addition		No. of Samples	Total Kjeldahl Nitrogen[1] ppm Mean	Range	S.D.	Nitrate Nitrogen ppm Mean	Range	S.D.	Nitrite Nitrogen ppm Mean	Range	S.D.	Ammoniacal Nitrogen ppm Mean	Range	S.D.	Other Components ppm
Ammonium Chloride Fig. 15	Ex	7	Y=298.8 - 3.753X, r = -0.63			0.1	0.1		0.04	0.06		Y=44.9 - 1.721X, r = -0.874			Organic nitrogen Y=254.8 - 2.886X, r = -0.615
	C	7	Y=257.8 - 3.753X, r = -0.63			0.1	0.1		0.04	0.05		2.2	1.5		
Lactose Fig. 16	Ex	6	Y=199.3 - 2.92X, r = -0.89			0.15	0.1		0.03	0.03		Cloudy interference			Lactose Y=249.1 - 47.8X, r = -0.83
	C	6	189	14		0.15	0.1		0.04	0.02					
Casein Fig. 17	Ex	7	310	35		0.2	0.1		0.03	0.07		Y= -0.6 + 5.23X, r = 0.965			Organic Dissolved(Ex) Y=104.5 - 7.91X, r = -0.90
	C	4	210	15		0.1	0.0		0.00	0.00		0.4	1.1		
Sodium Nitrite	Ex	6	189	31		0.4	0.6		31.92	5.00	2.14	3.4	2.8		
	C	6	188	28		0.1	0.1		0.11	0.05		3.3	1.9		
Sodium Nitrate	Ex	6	207	35	13.43	36.5	7.2	2.9	0.04	0.06		1.3	1.5		
	C	6	180	29	12.26	0.1	0.1		0.03	0.01		1.4	0.6		
Lact-albumin Fig. 18	Ex	6	318	20		0.1	0.1		0.05	0.04		Y= -0.3 + 6.62X, r = 0.92			Organic Nitrogen(Ex) Y=324 - 8.95X, r = -0.91
	C	6	207	11		0.1	0.2		0.02	0.03		1.8	0.1		Dissolved Kjeldahl(Ex) Y=99.1-4.34X, r = -0.63
Nitrate-Ammonia	Ex	5	194	8	3.64	22.2	3.8		0.15	0.26		Y=25.5 - 0.742X, r = -0.96[2]			
	C	3	177	1		0.1	0.2		0.03	0.00		4.0	1.3		
Nitrite-Ammonia	Ex	5	197	22	9.69	6.8	2.2		10.50	3.75		Y=25.5 - 0.742X, r = -0.96[2]			
	C	3	177	1		0.1	0.2		0.03	0.00		4.0	1.3		
Nitrite-Nitrate	Ex	6	203	10	4.29	32.0	7.1		9.9	3.20		4.9	1.1		
	C	3	177	1		0.1	0.2		0.03	0.00		4.0	1.3		
Anaerobic															
Nitrite-Nitrate Fig. 20	Ex	6	204	14	5.10	Y=32.5 - 2.17X, r = -0.96			Y=7.5 + 1.33X, r = 0.7			4.5	1.3		
Nitrite-Ammonia Fig. 19	Ex	5	Y=222 - 3.863X, r = -0.82[3]			7.4	10.5		11.15	4.42		25.5	1.9		
Nitrate-Ammonia Fig. 19	Ex	5	Y=222 - 3.863X, r = -0.82[3]			Y=31.6 - 2.34X, r = -0.99			Y=0.44 + 1.17X, r = 0.99			24.8	0.8		
	C	3	197	14		0.1	0.1		0.03		0.01	4.6	2.4		

Addition		pH Init.	pH Final	eH Init.	eH Final
Ammonium Chloride	Ex	6.8	6.9	+ 10	+ 8
	C	6.8	7.2	+ 10	- 10
Lactose	Ex	6.7	6.9	+ 20	+ 15
	C	6.8	7.1	+ 20	0
Casein	Ex	6.6	7.3	+ 25	- 15
	C	6.7	6.9	+ 20	0
Sodium Nitrite	Ex	6.8	7.2	+ 20	- 10
	C	6.7	7.3	+ 20	- 15
Sodium Nitrate	Ex	6.8	7.3	+ 20	- 15
	C	6.8	7.2	+ 20	- 10
Lact-albumin	Ex	7.2	8.2	- 5	- 65[4]
	C	6.8	7.5	+ 10	- 25
Nitrate	Ex	6.9	7.1	+ 15	0
Ammonia	C	6.9	7.4	+ 15	- 15
Nitrite	Ex	6.9	7.3	+ 10	- 15
Ammonia	C	6.9	7.4	+ 15	- 15
Nitrite-Nitrate	Ex	7.1	7.4	0	- 20
	C	6.9	7.4	+ 15	- 15
Anaerobic					
Nitrite-Nitrate	Ex	7.1	6.3	0	+ 5
Nitrite-Ammonia	Ex	7.0	6.3	+ 10	+ 45
Nitrate-Ammonia	Ex	6.9	6.3	+ 10	+ 45
	C	7.0	6.4	0	+ 40

1. Total Kjeldahl Nitrogen is equivalent to total nitrogen except where nitrate and nitrite are present in quantity.

2, 3. Equation based on combined results, as separation was not valid.

4. After 24 hours.

Y = Parts per million of component.

X = Time since start of experiment, hours.

Table 17.

Summary of Nitrogen Changes.

Aerobic Experiment	Nitrogen Level	
Addition	Experimental	Control
$NO_3 - NH_3$	NH_3 -N decrease	---
$NO_2 - NH_3$	NH_3 -N decrease	---
$NO_2 - NO_3$	---	---
NO_2	---	---
NO_3	---	---
NH_3	Total Kjeldahl/ decrease	decrease
	Organic decrease	decrease
Casein	NH_3 - N increase	---
	Dissolved organic decrease	---
Lactalbumin	Organic N decrease	---
	Total dissolved N decrease	---
	NH_3 -N increase	
Anaerobic		
$NO_2 - NH_3$ }	Total Kjeldahl decrease	
$NO_3 - NH_3$ }		
$NO_2 - NO_3$	NO_3 - N decrease	---
	NO_2 - N increase	---
	NO_3 decrease	---
	NO_2 increase	---

Fig. 15
AMMONIA ADDITION

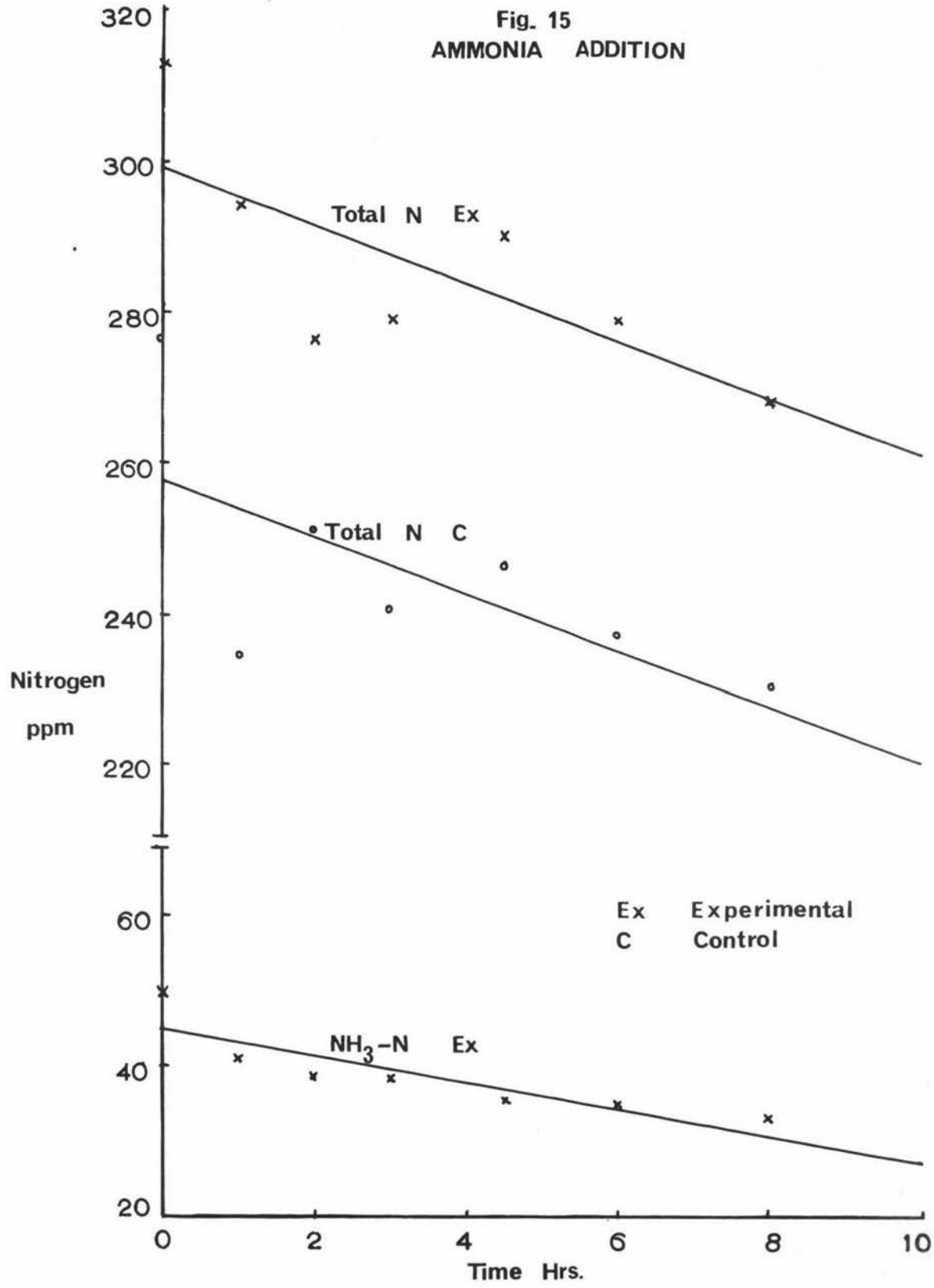
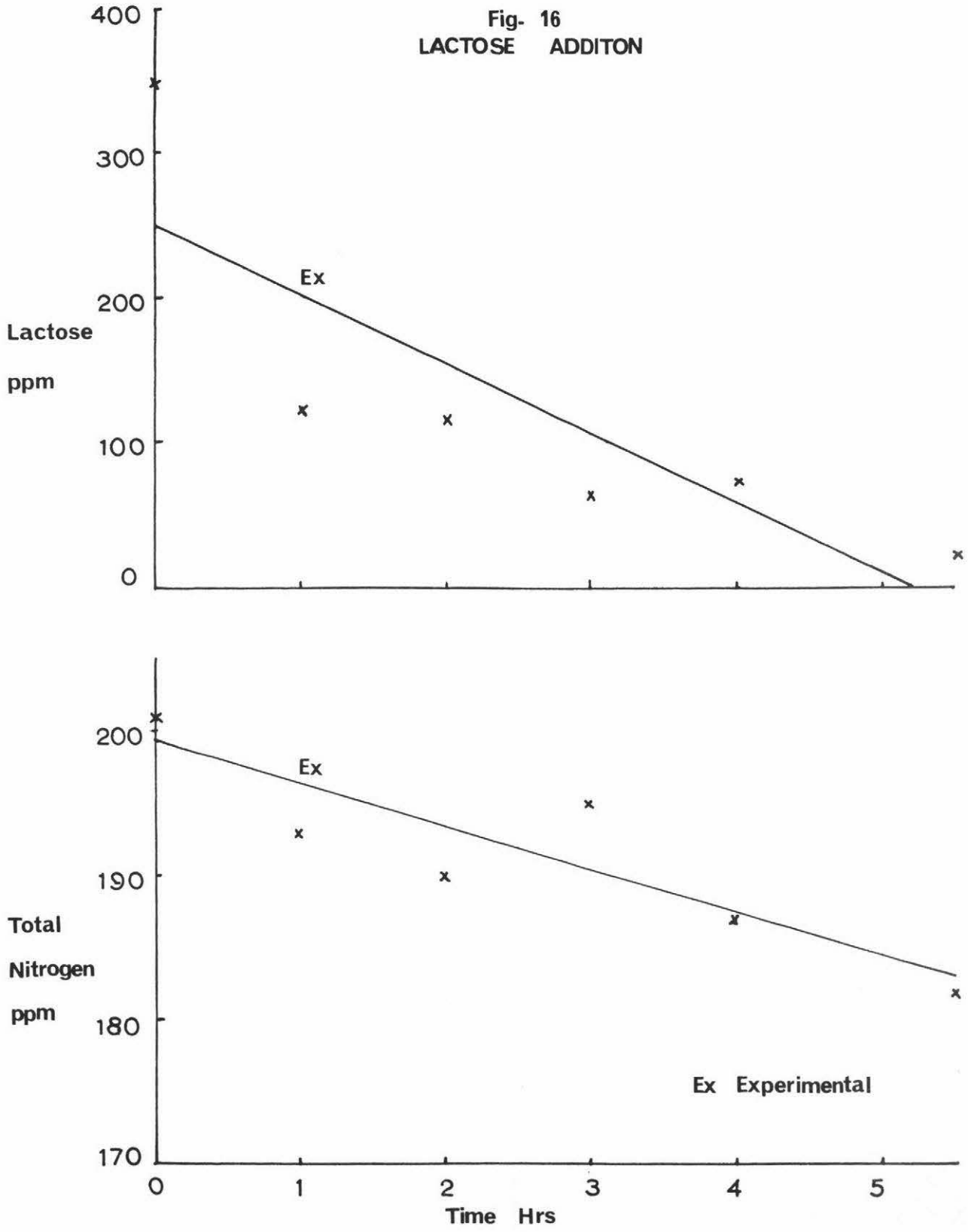


Fig. 16
LACTOSE ADDITON



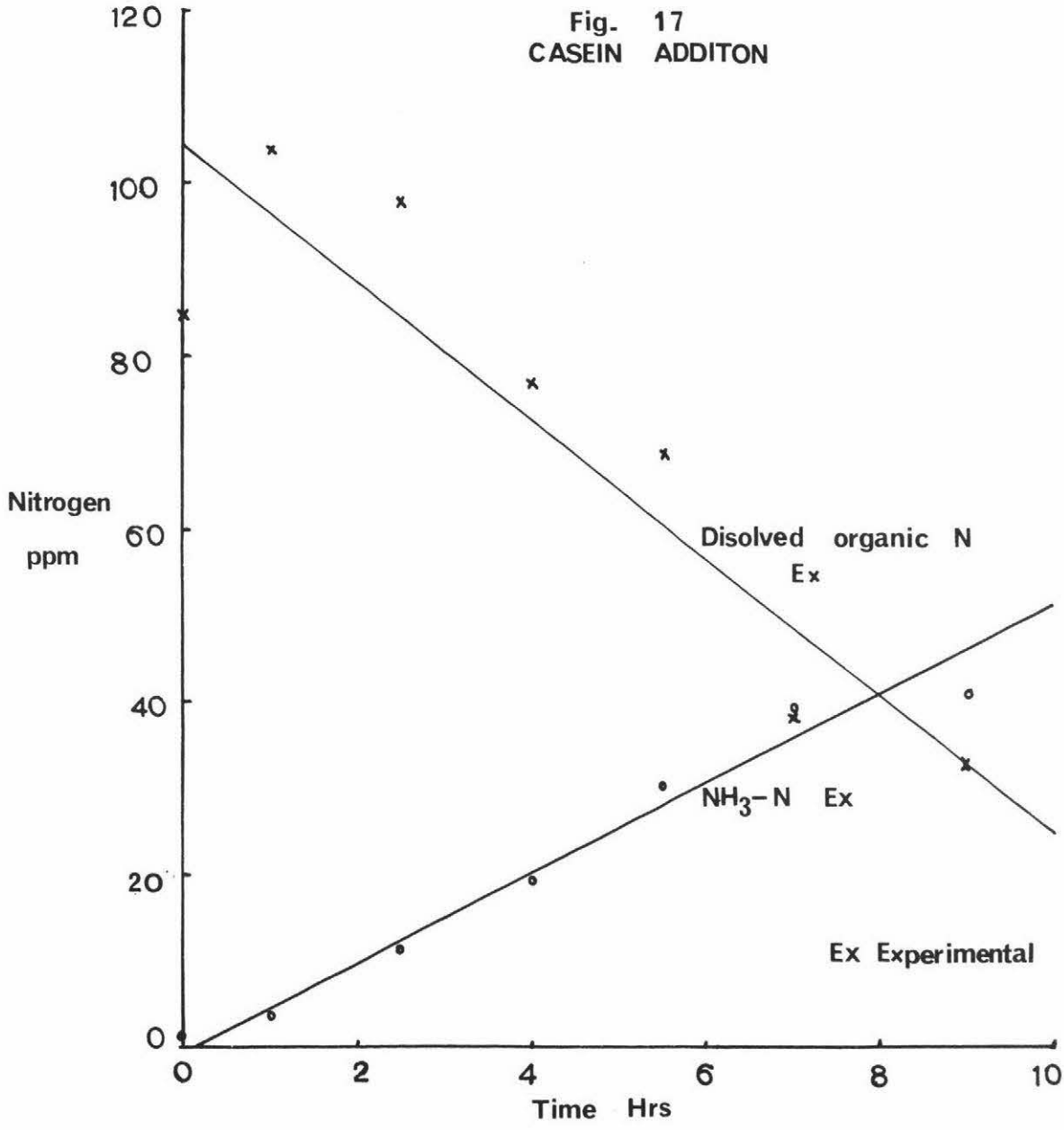


Fig. 17
CASEIN ADDITON
120
100
80
60
40
20
0
Nitrogen
ppm
Disolved organic N
Ex
NH3-N Ex
Ex Experimental
0
2
4
6
8
10
Time Hrs

Fig. 18

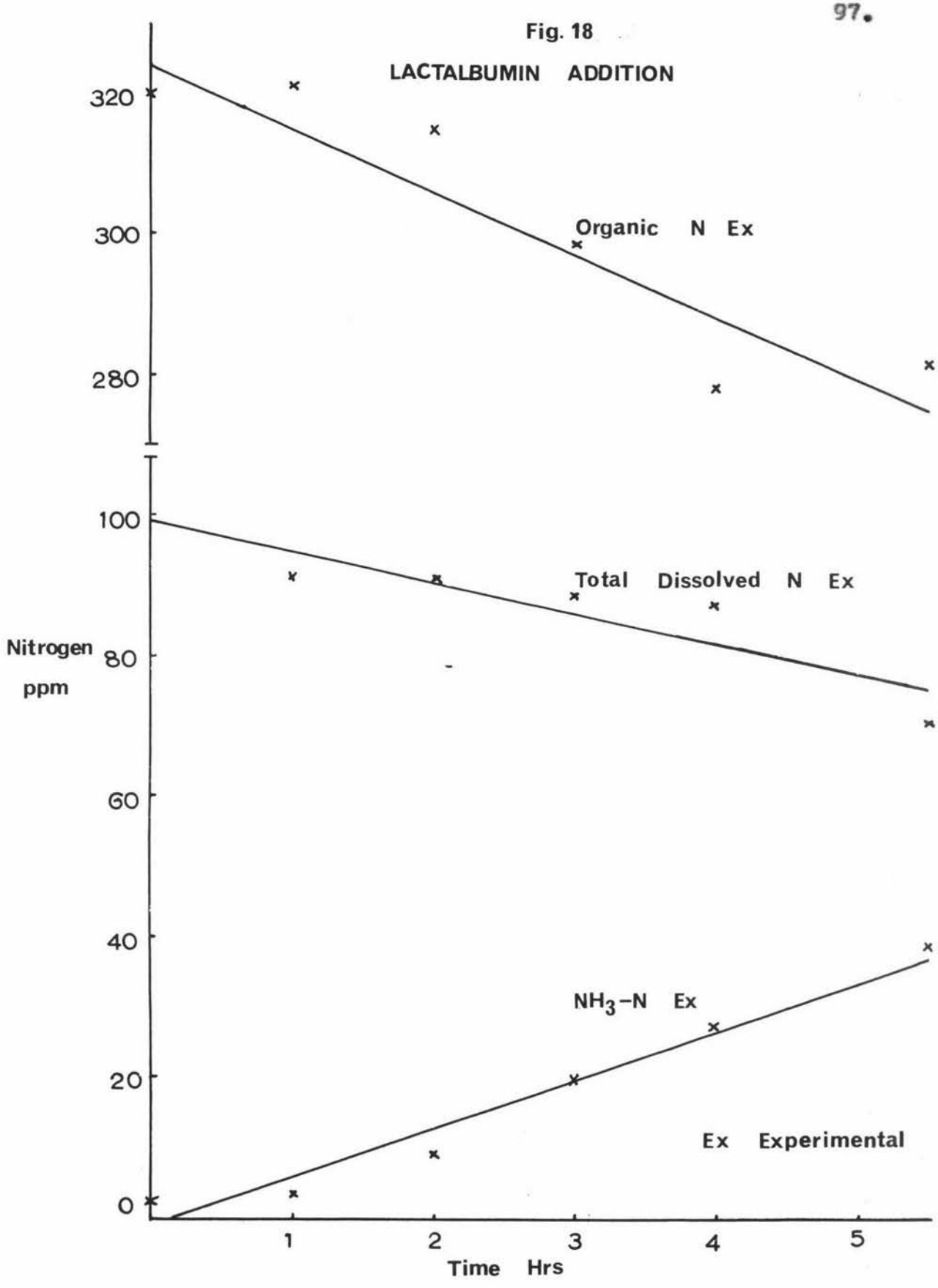

LACTALBUMIN ADDITION
320
300
280
Organic N Ex
100
80
60
40
20
0
Total Dissolved N Ex
Nitrogen ppm
NH3-N Ex
Ex Experimental
1
2
3
4
5
Time Hrs

Fig. 19

AMMONIA NITRATE AND

AMMONIA NITRITE ADDITION

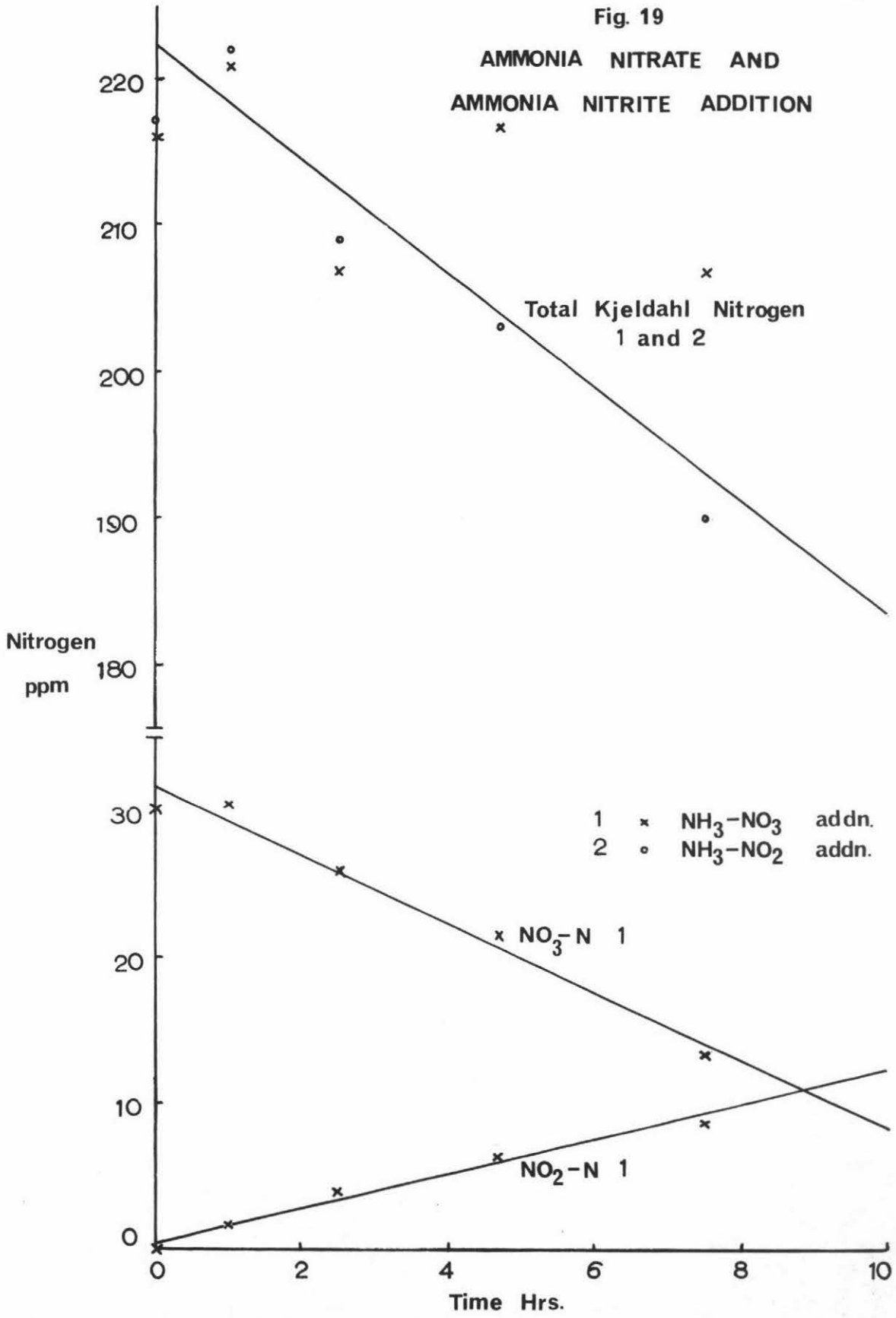
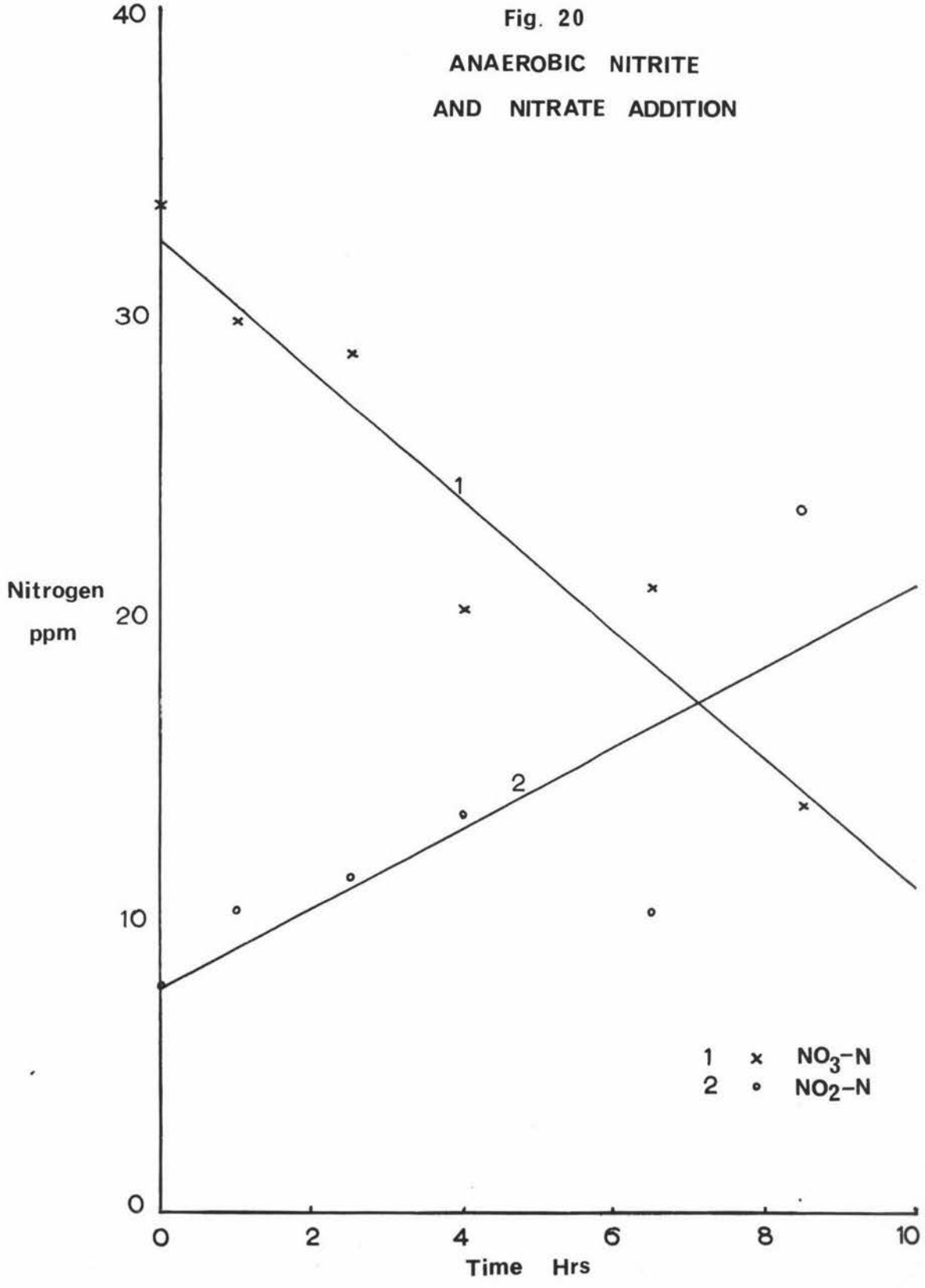


Fig. 20

ANAEROBIC NITRITE

AND NITRATE ADDITION



Discussion

Because of the limited number of samples that could be analysed from each experiment, in many cases a valid regression was not found. This was basically because of the considerable variability found with the small numbers of samples, and because the tests for statistical validity require a very high degree of correlation for small sample numbers. Despite this, the results collected provide valuable information.

Consideration of the initial and final pH levels and Redox potentials in Table 16 reveals that, although the pH levels were consistently near neutral, in all the aerobic experiments, the pH rose during the experiment, while the redox potential fell. The reverse occurred in the anaerobic experiments. This indicates the conditions were definitely different, although strictly anaerobic conditions would be unlikely to exist in the stationary flasks. The rise in the pH level in the lactose addition experiment indicates conditions were aerobic and the lactose was not converted to lactic and butyric acids.

There is no evidence of nitrification occurring at any stage. On the basis of the reports in the literature on methods for determining nitrifying ability, the aerobic experiments with ammonia added would have shown nitrate production if nitrifying organisms were present.

There is definite evidence of an overall loss of nitrogen in two aerobic cases, the aerobic ammonium chloride and lactose addition experiments. The lactalbumin addition experiment also provides nitrogen loss evidence, as the organic nitrogen level was decreasing more rapidly than the ammonia level was increasing, and the other forms of nitrogen changed only slightly. This

nitrogen loss confirms the observations of Leclerc and others that a nitrogen loss does occur from trickling filters. The question of mechanism for this nitrogen loss is not solved, however. The mechanism proposed by Leclerc requires the presence of nitrifying organisms to provide nitrite. The action of nitrifying organisms was not found. Also, if a mechanism such as Leclerc's applied to the filter, the aerobic nitrite addition should have enhanced the nitrogen loss. This was not found, and the added nitrite did not change in a statistically significant manner. A possible mechanism is that also mentioned by Leclerc

$$4\ NH_3 + 3O_2 = 6\ H_2O + 2\ N_2$$

This requires the absence of nitrites and nitrates and is slow. It is a possibility in view of the evidence that no total nitrogen loss was validly found for the nitrite and nitrate addition experiments. This was probably because of the stringent statistical requirements, however, and provides only flimsy evidence for this reaction.

Conditions in the aerobic shake flasks may have been sufficiently anaerobic for anaerobic denitrification to occur. This was unlikely, however, as the results in the anaerobic experiments were quite different to the aerobic ones. The intermediate, nitrite, was not found. Nitrogen loss by ammonia volatilisation was unlikely, as the necessary high pH was not present.

The decrease in dissolved organic and total Kjeldahl nitrogen and the increase in ammonia nitrogen was expected, the protein being hydrolysed and converted to ammonia prior to use by the organisms. The decrease in ammonia, on ammonia addition, was also expected, the ammonia being utilised by the organisms.

There appears to have been no utilisation of nitrite or

nitrate in the aerobic experiments. Nitrate removal was expected on the basis of the pilot plant results. Possibly nitrogen was not in such demand in the shake flasks. Another possibility is that the removal of nitrate in the pilot plant was by absorbtion into the biomass rather than by a biochemical transformation.

The anaerobic experiments gave results that were expected. Denitrification occurred definitely in two cases. Production of nitrite from nitrate was also found and this further indicates denitrification along the normal pathways was occurring.

Conclusions

(1) There is evidence that a loss of total nitrogen occurred when the organisms from the pilot plant were incubated under aerobic conditions. This supports the prediction of the literature.

(2) There is no evidence of the presence of nitrifying organisms among those collected and incubated. This does not support the mechanism proposed by Leclerc for the loss of nitrogen from trickling filters. A possible mechanism for the loss of nitrogen is

$$4\ NH_3 + 3O_2 = 6\ H_2O + 2\ N_2$$

(3) The collected organisms appeared to be able to perform denitrification along normal routes.

CHAPTER 4.

SOME OTHER ASPECTS OF THE EXPERIMENTAL UNITS.

INTRODUCTION.

Several aspects of the experimental plants are considered in this chapter. They are not necessarily related to each other, but provide further information on the plants. The aspects considered are lactose removal in the laboratory plant, oxygen transfer in the pilot plant, and the composition, digestion and preservation of the filter biomass.

LACTOSE REMOVAL IN THE LABORATORY SCALE PLANT.

This plant was operated on a batch-feed basis, and hence could be used for determination of the rate of lactose removal. The method used was simply the collection of samples from the plant on feeding and at known time intervals afterwards, and the analysis of these samples for lactose content. Three sets of results were collected. For each a statistically valid linear regression equation for lactose remaining versus time was computed. For the third set of data determination of the levels of the different forms of nitrogen was also made. The results are presented in Table 18.

As can be seen only two weeks after commissioning, the plant was capable of complete lactose removal within three hours. The rate of removal increased with the length of time the plant had been operating, the final rate being approximately three times the original. (The rate of removal is the coefficient of X in the equations of Table 18). This meant that, even at the much

Table 18.

Lactose and Nitrogen Removal

Period since Commissioning	Lactose Remaining Equation	(Y = ppm Lactose, X = time, hours Y_1 = ppm Organic Nitrogen)	
2 weeks	Y = 185.5 - 63.7 X	Complete removal 3 hours	
3 months	Y = 139 - 87.1 X	Complete removal 1¾ hours	
15 months	Y = 422 - 208 X	Complete removal 2 hours	
	Organic Nitrogen: Y_1 = 14.2 - 1.83 X	Complete removal 8 hours	
	Nitrate Nitrogen :	Mean 0.23 ppm	Range 0.1 ppm
	Nitrite Nitrogen :	Mean 0.01 ppm	Range o.04 ppm
	Ammoniacal Nitrogen :	Mean 3.6 ppm	Range 3.2 ppm

higher lactose level of the feed in the third set of data, total lactose removal was completed within approximately two hours.

Nitrogen removal was accomplished more slowly, eight hours being required for removal of the organic nitrogen. There was no evidence of nitrification occurring. Nitrification was not expected at the high hydraulic and organic loadings used.

The rapid removal of lactose and nitrogen, coupled with the high rates of BOD removal (97% plus) in the plant would suggest that it could have been operated at higher organic loadings.

This was not feasible, however, as at the higher organic loading used during the latter part of the plant operation, column blockage because of excess growth became much more frequent. Any higher organic loading would have made operation impractical.

OXYGEN TRANSFER IN THE PILOT PLANT

Prior to commissioning the pilot plant on its whey feed, an experiment was conducted to determine the rate of oxygen transfer from the atmosphere to the liquid passing through the column.

Theory

The sulphite oxidation method was used. This is based on the instantaneous oxidation by oxygen of sulphite ions, in the presence of a catalyst, to sulphate. The amount of sulphite oxidised is proportional to the oxygen transferred to the solution. It may be estimated by measuring sulphite levels before and after oxygen transfer, by reacting the sulphite solution with excess iodine, and measuring the residual iodine after reduction by the sulphite, by titration with thiosulphate solution.

By determining the amount of oxygen transferred to the solution after a single passage through the column, and by determination of the flow rate of the liquid through the column, the time rate of transfer of oxygen to the liquid could be calculated.

An oxygen transfer coefficient to a liquid may be defined by

$$K_L a = \frac{N}{V\,(C^* - C)}$$

where $K_L a$ = mass transfer coefficient, hour $^{-1}$

N = oxygen transfer rate, lb mole/hr

V = weight of water, lb mole

C* = equilibrium concentration of oxygen, lb mole/lb mole water

C = concentration of oxygen in solution, lb mole/lb mole water

C is zero in experiments using the sulphite method, as the dissolved oxygen is instantaneously combined to form sulphate.

Experimental

Three gallons of a 10,000 ppm sulphite ion solution were prepared, cobalt chloride at 3 ppm also being added. Samples were collected for sulphite determination before and after a single passage through the column. The pump used was the original diaphragm pump used for the plant, and this had a flow rate of one Imperial gallon per minute.

Results

The sulphite concentration was reduced by 6240 ppm on passage through the plant. This corresponded to an oxygen transfer of 791 ppm.

On the basis of a liquid flow rate of one gallon/minute, the rate of oxygen transfer was 0.0149 lb mole/hr.

C* for the temperature of 21°C. was calculated as

5.18×10^6 lb moles/lb mole water.

The weight of water aerated in one hour would have been 33.4 lb moles at the flow rate of 1 gallon/minute.

Therefore

$$K_L a = \frac{0.0149}{33.4 \times 5.18 \times 10^6}$$

$$= 86 \text{ hr}^{-1}$$

Discussion

This mass transfer coefficient compared quite favourably with values which had previously been determined for 14 litre laboratory fermenter, ranging from 54 hr^{-1} to 414 hr^{-1}, according to operating conditions. The pilot plant filter column was therefore a reasonably effective aeration device.

BIOLOGICAL COMPOSITION OF THE BIOMASS OF THE FILTERS.

Detailed examination of the filters' biomass was not undertaken. However, microscope slides of the biomass of the plants were prepared and Gram stained at intervals. The morphology of the major types of microorganism was noted.

Gram positive micrococci were the predominant form found in the laboratory plant. Also present in large numbers were coryneform bacteria, which were bipolar staining rods. Several unknown types of protozoa were also observed.

In the pilot plant, a progression in the predominant microorganisms was observed. In the first few months of operation, the predominant type were Gram positive micrococci, with coryneform bacteria present in considerable numbers. Gram-positive diplococci were also present.

After six months' operation the coryneform bacteria were more common, as may be seen in Plate 5. Micrococci were present in fewer numbers. Some branch-like pieces characteristic of the genus. Streptomyces were also observed. Plate 6 shows another

PLATE 5.

BIPOLAR STAINING CELLS, CORYNEFORM TYPE.

(MICROSCOPE MAG. 1000X)

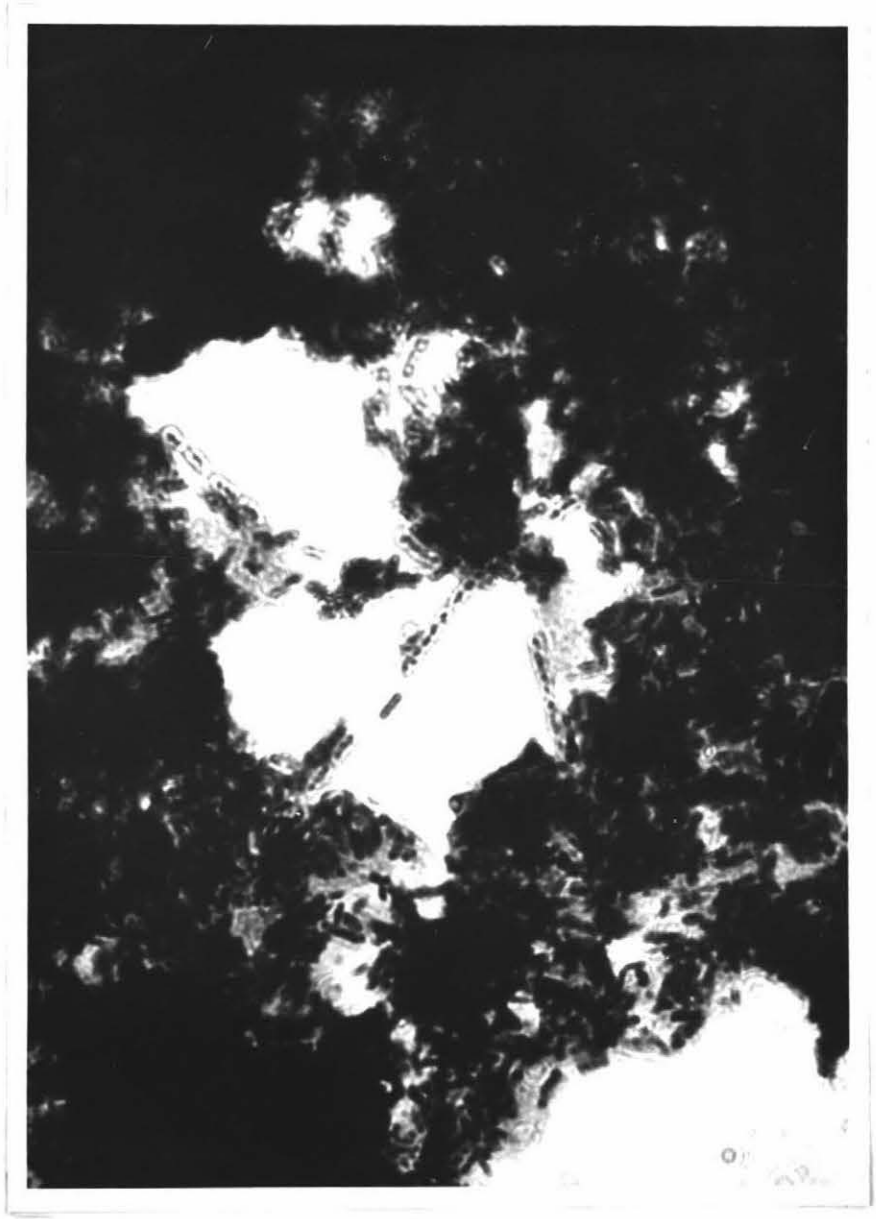


PLATE 6.

CELLS WITH NEGATIVELY-STAINING WALLS.

(MICROSCOPE MAG. 1000X)

type found, which formed long chains with a negatively-stained cell wall.

Small flies inhabited the pilot filter throughout its operation. These were not identified, but were presumably the typical filter fly of the genus Psychoda.

CHEMICAL COMPOSITION OF THE BIOMASS

The chemical composition of the biomass of the pilot filter in terms of moisture and protein content was determined concurrently with the shake flask experiments. Samples of the centrifuged sludge were analysed for moisture content (Appendix 1) and their protein content was calculated on the basis of the organic nitrogen content of the first samples taken from the shake flasks. The results are presented in Table 19. The protein level of the dried sludge indicates its suitability for use as a stock food or fertiliser.

Table 19.

Composition of the Biomass.

Total Solids %	Protein % (of dryweight)
11.0	35.8
9.5	43.9
9.5	31.5
8.1	41.5
6.8	43.8
10.3	33.6
7.5	39.9
Mean 9.0 Range 4.2	Mean 38.3 Range 12.4

The bacterial concentration in the sludge was estimated by preparing dilutions of the centrifuged sludge and carrying out a plate count using tryptone-glucose-yeast extract agar and incubating the plates for three days at 20°C. This temperature was used as it was the one at which the column was operated.

The bacterial concentration was found to be 2.41×10^{10} organisms/gram wet sludge, or 3.94×10^{11} organisms/gram dry sludge.

The total sludge for twenty-four hours operation was collected and centrifuged on one occasion. The sludge production was 545 gms of 'wet' sludge at 6.3% total solids, or 34.4 gms on a dry basis. This converts to 1.355×10^{13} organisms per day on the basis of the estimated bacterial concentration.

SLUDGE DIGESTION

A simple experiment was carried out to determine whether the sludge recovered from the pilot plant by centrifugation could be digested by anaerobic means.

Sludge from the anaerobic digestors of the local municipal sewage works was collected in an insulated vessel. Three bottles of 600 c.c. capacity were filled with 500 gms of the following:-

Bottle 1 - sludge from the digestors.

Bottle 2 - sludge from the digestors, plus 5% by weight of centrifuged pilot plant sludge.

Bottle 3 - centrifuged pilot plant sludge.

These bottles were incubated in a room at 37°C, the gas from the bottles being collected over water. The results are presented in Table 20.

Table 20.

Sludge Digestion Experiments

Days Since Start-up	Anaerobic Sludge	Anaerobic and Pilot Plant Sludge	Pilot Plant Sludge
1	150 cc gas burnt	300 cc gas burnt off	Slight gas production
3	No gas	300 cc gas off, 25 gm sludge added	No Gas
7	No gas	200 cc gas off	No Gas
10	No gas	150 cc gas off	No Gas
14	No gas	300 cc gas off 25 gm sludge added	No Gas
20	No gas	No gas	No Gas

The anaerobic sludge appeared able to digest the pilot plant sludge for a period of two weeks, and then became no longer active. Possibly centrifuged sludge was fed too frequently, or more quickly than the anaerobic flora could adapt to it. The conditions in the bottles were probably not ideal for the anaerobic organisms, and this possibly caused them to die off. It would appear that the centrifuged sludge could be successfully digested by anaerobic means if a properly designed digestor and an adapted flora were used.

PRESERVATION OF THE FILTER BIOMASS

The biomass from the pilot plant was preserved for future use in the following manner. Centrifuged sludge from the recirculation vessel was mixed to a pasty constituency with a large number of materials, which are listed in Table 21. The pastes were dried at ambient temperature under a vacuum of 25" Hg for five days, and

were then stored in a deep-freeze. It was hoped that, by using a wide range of materials, at least some would support a viable culture.

Table 21.

Biomass Preservation Materials

Sterile city soil
Sterile soil from outside the laboratory
Sterile soil from a park
Sterile soil from the University gardens
Clay
Filter aid (diatomaceous earth)
Glass beads
Polystyrene pieces
String
Starch
Fullers earth
Bentonite
Sand
Cellophane
Dried by itself
Frozen under paraffin
Frozen by itself

PLANT SIZE FOR A TYPICAL FACTORY

The size of tricking filter installation required to treat a waste flow to 40,000 gal/day from a cheese factory was calculated. The basis of the calculation was a direct scale-up of the pilot plant.

Waste Composition

It was assumed the waste had a typical composition of 1000 ppm BOD. For the filter to run on a continuous basis, balancing tanks would be required to even out the waste flow to the filter. The size of these would depend on the operations of the individual factory, and was therefore not calculated.

Filter Column Size

The pilot plant was able to provide a BOD reduction of

83% at an organic loading intensity of 1.5 lb BOD/yd^3day, at a recirculation ratio of 51.1 : 1. This ratio was near the optimum found for the plant, and the BOD removal and loading were in the range for a roughing treatment. A BOD loading of 1.5 lb/yd^3day was therefore chosen.

$$\text{The BOD loading on the plant} = \frac{40000 \times 10 \times 10^3}{10^6}$$

$$= 400 \text{ lb/day}$$

$$\text{Therefore, the filter volume} = 1.5 \times 400$$

$$= 600 \text{ yd}^3$$

For a circular filter of depth eight feet,

$$\text{Diameter} = \sqrt{\frac{600 \times 27 \times 4}{8 \times \pi}}$$

$$= 51 \text{ ft}$$

Recirculation Tank

On the basis of the pilot plant tank of 25 gallons for a waste flow of 170 gallons, the required volume would be 5880 gallons. The tank size in the pilot plant was used because it was the only one available, and the ratio of its size to waste flow was arbitrary. A smaller tank could probably be used.

RECIRCULATING PUMP SIZE.

For the optimum recirculation ratio of 45 : 1, a pump of capacity calculated below would be required.

$$\text{Pump capacity} = 46 \times \frac{40{,}000}{24}$$

$$= 76{,}600 \text{ gallons/hour.}$$

Settling Tank Size

For a sedimentation time of one hour, a volume of 1,666 gallons would be required.

The table below (Table 22) summarises the size of plant required.

Table 22.

Plant to Treat 40,000 gallons/day of Cheese Factory Waste.

Filter Column	8' depth, 51' diameter, 600 yd^3 capacity
Pump	Capacity 76,600 gal/hr.
Settling Tank	Capacity 1,666 gal.

CONCLUSIONS

(1) Lactose removal in the laboratory plant was very rapid and generally complete within two hours. Organic nitrogen removal required eight hours.

(2) The pilot plant column was an effective oxygen transfer device, with an oxygen mass transfer coefficient of 86 hr^{-1}.

(3) The most common microorganisms found in the plants were coryneform bacteria and micrococci, on a morphological basis.

(4) The centrifuged sludge or biomass from the pilot plant had a typical composition of 9% total solids and 3.4% protein. When dried, its protein content was typically 38%.

(5) The sludge would be amenable to anaerobic digestion in a suitably designed plant.

CONCLUSIONS AND SUGGESTED FURTHER RESEARCH.

A pilot-scale trickling filter capable of providing a satisfactory 'roughing' treatment for a compounded dairy factory waste was constructed. The plant was designed on a conventional basis, but was operated at high hydraulic and organic loading intensities. BOD removals of 60 - 85% were typically obtained.

The controversial field of theoretical and empirical performance prediction relationships for trickling filters was not resolved. The experimental data collected fitted a curve described by the equation

$$Y = 17.778 + 3.079X - 0.0342X^2$$

where Y = % BOD removed by the filter

X = recirculation ratio

This equation predicts an optimum BOD removal at a recirculation ratio of 45 : 1, which is considerably higher than the 10 : 1 ratio commonly used. The equation is specific to the pilot plant. Further experimental work at more closely controlled hydraulic and organic loading intensities would be required to determine if the predictions of the equation have wider application.

The pilot plant operated satisfactorily at BOD : N ratios in the raw waste in the range 21 - 27 : 1. These are higher than the generally recommended maximum of 20 : 1. Despite this, typically 30% of the nitrogen in the raw waste was found in the plant effluent. There was no evidence of nitrification in the pilot plant. Further study of plant performance at a range of BOD : N ratios would clarify whether the recommended ratio applied to this type of plant.

Study of the nitrogen balance of biomass from the plant incubated in aqueous suspension indicated that a net loss of nitrogen occurred under both aerobic and anaerobic conditions. Denitrification under anaerobic conditions appeared to follow normal pathways, but the mechanism for the aerobic nitrogen loss was not determined. There was no evidence of the presence of nitrifying organisms, which could possibly be responsible for nitrogen loss under aerobic conditions, according to Leclerc (1959). Further research using greater sample numbers would provide more information on possible mechanisms for this nitrogen loss.

APPENDIX 1

ANALYTICAL METHODS

Lactose Content

This was estimated using the volumetric semi-micro method of Somogyi (1945). This is based on the use of an alkaline copper tartrate reagent.

Biochemical Oxygen Demand (BOD)

The procedure given by the American Public Health Association, the American Water Works Association and the Water Pollution Control Federation (1960) in Standard Methods for the Examination of Water and Wastewater including Bottom Sediments and Sludges, page 309 was followed. Seeding of the samples was found to be necessary. Oxygen content was determined using the azide modification of the Winkler method for dissolved oxygen. Dilutions containing 0.25% of the plant feeds, 1% of the pilot plant effluent and 10% of the laboratory plant effluent gave satisfactory oxygen reductions. Samples were settled for thirty minutes prior to dilution.

NITROGEN DETERMINATIONS

As recommended by A.P.H.A., A.W.W.A. and W.P.C.F. (1960) the samples were preserved for up to twenty-four hours with 1500 ppm sulphuric acid and stored at 20°C.

Total Kjeldahl Nitrogen

The basic method used was that of the A.P.H.A. et al (1960), page 307. For most of the experiments the distillation step was modified to a semi-micro method. Better reproducibility

of results was obtained by this method.

Nitrate Nitrogen

This was determined by the Brucine method, as described by the A.P.H.A. et al (1960), page 178. Nitrite interference was encountered in the nitrite addition experiments. This was overcome by prior measurement of the nitrite, followed by oxidation of nitrite to nitrate using acidified potassium permanganate. The total nitrate present was then determined and the original nitrate calculated as the difference between this and the nitrite concentration.

Nitrite Nitrogen

The method given by the A.P.H.A. et al (1960), page 303 was used. This involved the use of sulphanilic acid and naphthylamine hydrochloride. The potential carcinogens in this latter material necessitated especial care in its handling. Prior clarification of the samples by centrifugation was required.

Ammoniacal Nitrogen

The direct Nesslerisation method described by the A.P.H.A. et al (1960), page 296 was used. Clarification by centrifugation was required.

Colourimetric Equipment

For nitrite, nitrate and ammonia, an Hitachi Model 101 spectrophotometer and cells with a 1 cm light patch were used.

Total Solids Content

This was determined on the centrifuged sludge by weighing

10 gms of sludge into a dried, tared and lidded moisture dish and drying overnight at 102°C.

pH and eH

These were measured on a Metrohm pH meter.

APPENDIX 2

PILOT PLANT DAILY LOG SUMMARY.

WEEK NUMBER

1 The plant was commissioned using the large 245 gallon settling tank. Within three days a brown-fawn film began to develop on the column. Because the pH dropped to 5.2, the diluted whey feed of 157 gal/day was shut off and an air pump used to aerate the settling tank. The stirrer in the tank was also turned on.

2 The pH began to rise because of the effect of the air pump. The whey feed was restarted at half the original rate. The pH rose to 6.8 by the end of the week, and the unpleasant odour became less noticeable.

3 The liquor in the settling tank began to turn brown. Small flies were noticed on the column. Growth in the return line from the column caused a blockage which required clearing. The pH rose to 7.1

4 Floc formation became noticeable in the settling tank. Maggot-like larvae were also observed. Because the pH had risen to 7.7, the air pump was switched off. The diaphragm recirculation pump failed and was replaced with a gear pump from a small filter press, with a flow rate of 2.66 gal/min compared with the 1 gal/min of the diaphragm pump.

5 The air pump was switched on again for one day, as the pH

WEEK
NUMBER

had dropped below 6.0. The stirrer was switched off, and within one day of this approximately half the tank surface was covered with floating floc material.

6 The smell from the plant became objectionable. The liquor colour became a deep red. The water feed valve had become shut off and was replaced with an orifice from a hypodermic syringe.

7 Because the pH began to drop, the settling tank volume was reduced to 119 gallons in an attempt to increase the frequency of passage of the liquor through the column. This had little effect.

8 In view of the paper by Harding (1952), a new system of a small recirculating tank and a larger settling tank for the net outflow from the plant was installed. The original diaphragm pump was replaced into the system, with an increased flow rate of 2.5 gal/min. The plant now had a recirculation ratio of 40.6 :1. The pH soon began to rise.

9 The pH during this week was on average 6.5, with a range of 0.4.

10 A centrifugal pump replaced the diaphragm pump, and the flow rate was adjusted to 2.61 gal/min i.e. a recirculation ratio of 43 : 1, by means of a gate valve on the pump outlet.

WEEK
NUMBER

11 Considerable washout from the column occurred during this week. The whey feed had failed for one day, and possibly the organisms had died from starvation and had been washed out. The pH was typically 6.8.

12 An unpleasant odour developed on one day. The whey had a sweet smell at this time. Foaming became noticeable in the recirculation tank.

13 The small settling tank for the net waste flow was eliminated in an attempt to reduce the odour problem it sometimes caused.

14 The pump flow rate was dropped to half its previous rate, to give a recirculation ratio of 20.32 : 1.

15 The plant pH dropped to 6.0 at this reduced flow rate, but because there was no odour problem it was held at this rate.

16 There was considerable washout from the column during the week. The pH rose to an average of 6.4.

17 The small filter flies were particularly noticeable this week. The pump flow rate was increased to a recirculation ratio of 30.54 : 1.

18 The cheese whey supply ceased because of the close of the dairy season. A reconstituted whey was made up at a strength of 5.67% w/v. The whey powder used was spray

Week Number

dried rennet whey powder, with a typical analysis of 14% protein, 0.6% fat, 7.0% ash, 70% lactose, 2% iron, 0.5% phosphorus and 0.5% calcium. Initially, skim milk powder was incorporated into the feed, but precipitation of the protein in the feed tank led to its elimination. An insulated feed tank to hold the chilled whey concentrate was installed. The metering pump rate was slightly altered so the recirculation ratio became 28.8 : 1.

19 A plate heat exchanger coupled to a constant temperature water tank was fitted into the exit line of the recirculating pump. Growth of biomass caused this to block very rapidly. The pumped hot water system was connected instead to a twenty-foot copper coil which was placed in the recirculation tank. This gave good temperature control at 20°C.

20 Overflow of the water header tank for one night caused starvation of the biomass on the column and subsequent washout.

21 The pump flow rate was found to be decreasing, and the pump was opened up to full flow in an attempt to clean the lines, and was then reset at its previous rate.

22 There was very little washout from the column during this week.

WEEK
NUMBER

23 The pump flow rate was increased to a recirculation ratio 38.8 : 1. This caused a pH rise to typically 7.0. The pump lines required frequent plugging with a bottle brush in order to hold the flow rate.

24 The feed of whey concentrate failed for two days, and there was considerable washout. The plant soon recovered.

25 The flow rate was increased to a recirculation ratio of 51.5 : 1. Foaming became noticeable.

26 A constant head/flow switch device was fitted at the top of the column, the switch being connected to the pump. This was an attempt to control the flow rate despite blockage in the pump lines. The float switch level was adjusted to give a recirculation ratio of 54.5 : 1. The slight level variation in the float switch device caused a slight level variation in the recirculation vessel and this helped to flush out washout from the column.

27 There was heavy foaming in the plant. Despite the float switch, the pump lines and the pump required cleaning.

28 The pump starter mechanism failed and required scraping down to remove excess carbon which was caused by the frequent on/off switching.

29 Partially blocked pump lines caused the pump to run continuously. The pump lines were cleaned.

WEEK
NUMBER

30 The pump line again blocked up, and there was considerable foaming in the recirculation tank. The pump starter switch required cleaning.

31 On the failure again of the pump switch, the float switch device was removed from the system and the pump flow rate adjusted to a recirculation ratio of 54.5 : 1.

32 The pump line blocked off completely. This was the first time this had happened. The lines were cleared.

33 The plant was shut down at the end of this week, because the experimental work had been completed. For shut-down, the plant was run on water only for two days, then allowed to dry out, and the biomass was then washed out of the column. Neither the concrete column nor the stones showed any deterioration.

APPENDIX 3

MATHEMATICAL AND STATISTICAL METHODS

Calculation of the Parabola for BOD Removal Versus Recirculation Ratio

The method used was curvilinear regression analysis, as outlined by Ezekiel (1941). The means of the BOD removals at the different recirculation ratios were used in this analysis. The method involved the determination of the constants a, b and c in the generalised parabola,

$$Y = a \times bX = cX^2 .$$

These constants were determined using statistical terms for simultaneous equations in b and c.

Testing of Operating Equations for Trickling Filters Using Experimental Data.

The bulk of the operating equations required flow rate to be expressed in mgad (U.S.). Flow rates for the pilot plant were calculated as Ig/yd^3day, and were converted to mgad (U.S.) on the basis of the column diameter, depth and volume. To test the applicability of the equations, the predicted performance, in terms of feed BOD remaining in the effluent, was calculated and plotted versus flow rate. The actual performance of the plant was also plotted versus flow rate, and the curves compared.

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